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# FUEL QUALITY/PROCESSING STUDY

## VOLUME IV - ON SITE PROCESSING STUDIES

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GULF RESEARCH & DEVELOPMENT COMPANY - FINAL REPORT

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ON-SITE PROCESSING STUDIES

Prepared For


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Lewis Research Center  
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16. Abstract  <p>The final report for this study consists of four volumes. Volume I, Overview and Results, presents the conclusions the study obtained from its evaluation of the feasible paths from liquid fossil fuel sources to generated electricity. The segments from which these paths were built are the results from the fuel upgrading schemes, on-site treatments, and exhaust gas treatments detailed in the subsequent volumes.</p> <p>This volume, Volume IV, presents the results of the study related to treating the fuel at the turbine and processing the turbine exhaust gas at the turbine site. Fuel treatments protect the turbine from contaminants or impurities either in the upgrading fuel as-produced or picked up by the fuel during normal transportation. Exhaust gas treatments provided for the reduction of NO<sub>x</sub> and SO<sub>x</sub> to environmentally acceptable levels. Volume IV results also consider the impact of fuel quality upon turbine maintenance and deterioration. On-site costs include not only the fuel treatment costs as such, but also incremental costs incurred by the turbine operator if a turbine fuel of low quality is not acceptable upgraded.</p>					
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## Introduction

Costs have been generated for fuel washing/treatment operations and exhaust gas DeNOx for selected cases. These costs are presented as incremental costs over a base case.

(1) The base case for cost estimating purposes for the combined cycle is as follows:

- Combined Cycle Power Plant
- Nominal 400 MW plant output
- Maximum fuel flow 400 gpm
- Distillate fuel
  - no fuel treatment system required
  - exhaust gas DeNOx not required
- Exhaust flow =  $8.4 \times 10^6$  lb/hr.

(2) The base case for the simple cycle is as follows:

- Simple Cycle Power Plant
- 30 MW plant output
- Maximum fuel treatment system fuel flow is 33 gpm
- Distillate fuel
  - no fuel treatment system required

As noted above, the base case combined cycle operates on distillate fuel. Consequently, all costs (capital, operating and maintenance) associated with installation and operation of a fuel treatment system and the exhaust gas DeNOx system are, by definition, incremental over the base case. Similarly, costs for installation and operation of a fuel treatment system are incremental for the simple cycle case. No costs have been generated for exhaust gas DeNOx for the simple cycle case since this technology is not applicable at simple cycle exhaust gas temperatures.

Costs are presented for the following items:

- (1) power plant operating and maintenance costs (including hot gas path parts replacement costs, turbine cleaning costs)
  - simple and combined cycle cases, distillate and ash-forming residual fuels with as-burned fuel alkali levels of 0.5, 1.0 and 2.0 ppm sodium.
- (2) fuel treatment system capital, operating and maintenance costs for the combined cycle and simple cycle cases
  - for 50 ppm alkali residual fuel supplied to the power plant site.
  - as-burned fuel alkali levels of 0.5, 1.0, 2.0 ppm sodium.
  - provision for other sodium levels in the fuel supplied to the power plant site.
- (3) exhaust gas DeNOx system capital, operating and maintenance costs for the combined cycle case
  - for 450 ppm NOx level exhaust gas (corresponds to the maximum NOx expected at 2.0% nitrogen in the fuel).
  - 90% effectiveness, i.e., NOx reduction from 450 ppm to 45 ppm in the stack effluent.

The data in this report includes a complete cost estimate case for high nitrogen, ash-forming fuel in simple cycle and combined cycle applications.

Fuel treatment costs are presented for petroleum residual fuel. Coal-derived liquids and shale oils are not expected to require on-site treatment since CDL's are expected to be commercially available as essentially ash-free distillates, whereas shale oils produced by above ground retort will be upgraded at the conversion facility to reduce gums and meet transportation requirements. The upgrading is expected to reduce trace element constituents, e.g., arsenic



(whose corrosive effects are not known), nickel, vanadium, etc. to levels tolerable by today's gas turbines.

NO<sub>x</sub> reduction by exhaust gas treatment (catalytic DeNO<sub>x</sub>) is treated generically as a function of nitrogen content and DeNO<sub>x</sub> effectiveness. Costs at effectiveness levels of 77.5%, 85% and 90% are presented in this report.

SO<sub>x</sub> reduction by exhaust gas treatment by three processes (lime/limestone, Wellman-Lord, and Shell-UCP) has been evaluated. Economics are expressed in \$/KWHr cost for sulfur removal levels consistent with fuel sulfur levels from 0.8 to 2.5 weight percent sulfur in the fuel.

## 1.0 COSTS ASSOCIATED WITH ON-SITE FUEL CLEANUP/TREATMENT

### 1.1 POWER PLANT OPERATING AND MAINTENANCE COSTS

This cost reflects the expected cost for operating, periodic routine turbine inspections, turbine and compressor cleaning, maintenance of the combustion system, turbine controls, all ancillary components, and any replacement of prematurely deteriorated hot section components due to corrosive contaminants. It does not include the operating and maintenance costs associated with the fuel washing/treatment system.

These costs are highly variable as a consequence of type of turbine, duty cycles, preventative maintenance practices, labor costs and accounting procedures. Therefore, the costs are presented as incremental to the costs for operation and maintenance of a standard base case, i.e., a 400MW combined cycle burning distillate fuel having low enough nitrogen to meet EPA regulations on  $\text{NO}_x$  emissions without the use of exhaust gas  $\text{DeNO}_x$  cleanup.

Table 1.1-1 shows these costs in mills per kWh of net power generated, in 1980 dollars. Although these may not be the same for specific cases, they will be used in this study for making the comparisons of the incremental cost of each fuel, fuel cleanup, and exhaust gas cleanup. Several sources of information were used to arrive at these as an average of several types of gas turbines:

1. The 1978 costs reported by utilities to the Federal Power Commission<sup>(1)</sup> have been examined to determine the incremental costs between the simple cycle peaking service and the mid range combined cycle, burning clean distillate fuels ( $<.5\text{ppm}$ ) sodium. The range of costs for the simple cycle was 3.0 to 8.5 mills/kWh with an

average of 5 mills/kWh, in 1978 dollars. For the combined cycle, because of the larger energy production over which to distribute these costs and because there are fewer starts, these costs are found to be an average of 2.5 mills/kWh or half of the simple cycle, again in 1978 dollars.

2. Costs are augmented by the turbine cleaning required for operation with ash-forming fuels such as residual fuels. These costs do not include the operation and maintenance of the on-site fuel treatment system; this is covered separately in section 1.2. Based on experience burning both distillate and residual fuel in turbines of similar design, the cost of operation with ash-forming fuels is, on an average, 3.0 mills/kWh more than that of operation on distillate in the simple cycle and 2 mills/kWh more for the combined cycle.
3. The effect of sodium levels in fuels on the replacement frequency is estimated from experience. For a fuel containing 1ppm Na an additional 0.4 mills per kWh in replacement parts is expected. For a fuel containing 2ppm Na an additional 1.6 mills per kWh is expected. These costs are incremental to that for a fuel containing 0.5 ppm sodium.
4. Escalation of costs in 1978 to those of 1980 have been based on the compounding of 10% per year for two years.

Reference:

"1978 Gas Turbine Electric Plant Construction  
Cost and Annual Production Expenses".

U.S. Government Energy Information Administration.  
(To be Published).

TABLE 1.1-1

POWER PLANT OPERATING AND MAINTENANCE COSTS (MILLS/kWh)  
(incremental costs over a Base Case\*)

for Simple and Combined Cycles  
Burning either Distillate or Ash-Forming  
Fuels Containing Given Levels of Sodium

<u>Na in fuel ppm</u>	<u>Simple Cycle</u>		<u>Combined Cycle</u>	
	<u>Distillate</u>	<u>Ash Forming</u>	<u>Distillate</u>	<u>Ash Forming</u>
.5	3.1	6.1	Base*	2.0
1.0	3.5	6.5	.4	2.4
2.0	4.7	7.7	1.6	3.6

Base Case is 400 MW combined cycle burning #2 distillate  
meeting EPA emissions without exhaust gas DeNO<sub>x</sub>,  
running 3000- 7000 hrs/year.

## 1.2 ON SITE RESIDUAL FUEL TREATMENT COSTS

A residual fuel treatment system, capable of supplying a specified quality fuel to the selected 400MW combined cycle base load power plant, would have a maximum fuel flow rate capacity of 400 gpm. At today's state-of-the-art for residual fuel treatment, this probably represents the largest modular arrangement. Larger installations would use a number of such modules.

The cost of operating a fuel treatment system comprises the capital cost as well as operating and maintenance cost.

All of the costs shown for residual fuel treatment are over and above the cost of operation on distillate fuel, which does not require fuel treatment.

### 1.2.1 Fuel Treatment System Capital Cost

The 400 gpm (400MW equivalent) fuel treatment system costs have been derived for a reference petroleum residual fuel for three levels of sodium contamination remaining after washing, and for two different types of fuel washing hardware. The three sodium levels are those selected for the gas turbine maintenance cost evaluation. The two types of fuel washing hardware are the electrostatic coalescer desalter (ED) and the centrifuge desalter (CD).

Table 1.2.1-1 shows the significant properties of the reference fuel and the typical total installed costs of the two types of systems for the three output sodium levels.

In order to make these basic costs consistent with turbine operating costs, with which they are combined later in this report, the following cost parameters have been derived for the 400 MW power plant operating at full nameplate rating and at three utilization rates.

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An annual capital cost of 14.6% of the total installed capital cost has been used because it is the average being used by U.S. utilities.

CAPITAL COST

Sodium Output Level-ppm Type of Desalting	<u>0.5</u>		<u>1.0</u>		<u>2.0</u>	
	<u>ED</u>	<u>CD</u>	<u>ED</u>	<u>CD</u>	<u>ED</u>	<u>CD</u>
Total Installed Cost- $\$10^6$	7.8	9.3	6.4	7.5	5.7	6.6
* Annual Capital Cost- $\$10^6$	1.14	1.36	0.94	1.10	0.83	0.97
Cost- mills/kWh						
3000 hrs/yr	0.95	1.13	0.78	0.91	0.69	0.80
5000 " "	0.57	0.68	0.47	0.55	0.42	0.48
7000 " "	0.41	0.49	0.33	0.39	0.30	0.34

\* 14.6% of the total installed capital cost

Table 1.2.1-1

## RESIDUAL FUEL TREATMENT SYSTEM

## CAPITAL COST

A. System Description

Fuel Flow Rate -	- 400 gpm max.; 350 gpm typical
Fuel Specific Gravity -	0.96 max.
Fuel Viscosity 100°F -	900 cSt max.
Effluent Water Oil Content -	15 ppm max.
Input Sodium Level -	50 ppm max.

B. Typical Fuel Treatment System Costs

	0.5		1.0		2.0	
	ED	CD	ED	CD	ED	CD
Output Sodium Level, ppm						
Type of Desalting						
Total Equipment Cost-\$X10 <sup>6</sup>	5.8	7.3	4.4	5.5	3.7	4.6
Installation Cost-\$X10 <sup>6</sup>	2.0	2.0	2.0	2.0	2.0	2.0
Total Installed Cost-\$X10 <sup>6</sup>	7.8	9.5	6.4	7.5	5.7	6.6

C. Major Equipment Included in Treatment System

Packaged desalter system (ED=Electrostatic; CD=Centrifugal)  
 Packaged effluent water cleanup system  
 Inhibitor injection and mixing equipment  
 Residual fuel forwarding and filtration equipment  
 Fuel heaters  
 Fuel analyzer

Note:

The equipment costs shown are typical values, and are not to be construed as selling prices for any given application. Desalting systems are individually specified for a given application and can vary considerably in cost.

Installation cost at different sites with different systems will also vary from the typical cost shown.



#### 1.2.1.1 Fuels Outside of Reference Fuel Specifications

For fuels with specific gravity, viscosity and output sodium/input sodium values outside the reference fuel defined in Table 1.2.1-1, multiplication factors can be used for the total installed costs. The reference case with a 1.0 factor is the case shown in Table 1.2.1-1 for a reduction in sodium from 50 ppm in the raw fuel to 1ppm in the washed fuel.

The significant effect of fuel specific gravity, viscosity level and sodium reduction is in the fuel washing equipment. In a previous report, these relative capital costs were developed based only on this equipment. A more usable number is the relative total installed capital costs. Also further investigation has shown that within the accuracy of this study there is no significant difference in capital cost between a system for a fuel under 20 cSt viscosity at 60°F and a fuel between 20 cSt at 60°F and 900° cSt and 100°F. The cost of the steam type fuel heaters involved is a small fraction of the total treatment system installed cost.

The relative total installed capital costs for fuels other than the reference fuel in Table 1.2.1-1 are listed in Table 1.2.1-2.

Table 1.2.1-2

RELATIVE TOTAL CAPITAL COSTS OF ON-SITE  
FUEL TREATMENT SYSTEMS

<u>Fuel</u> <u>Viscosity</u>	<u>Fuel Specific</u> <u>Gravity</u>	<u>Sodium Out/</u> <u>Sodium In</u> <u>(ppm)</u>	<u>Total Capital</u> <u>Cost Factor</u>
Under 900cSt at 100°F	0.96 max. (or blended to 0.96 max.)	2/50	0.90
		1/50	1.00
		0.5/50	1.20
		2/20	0.90
		1/20	0.95
		0.5/20	1.10
		2/10	0.90
		1/10	0.95
		0.5/10	1.0
Over 900cSt at 100°F	0.97-0.985 (or blended to 0.97-0.985)	2/50	1.00
		1/50	1.20
		0.5/50	1.40
		2/20	1.00
		1/20	1.10
		0.5/20	1.30
		2/10	1.00
		1/10	1.10
		0.5/10	1.20

### 1.2.2 Fuel Treatment System Operating and Maintenance Cost

The major operating costs of a residual fuel treatment system can be summarized as:

A. Chemical Cost

1. Vanadium inhibitor
2. Demulsifier

B. Utilities Cost

1. Water for fuel washing
2. Electricity
3. Steam for fuel heating

C. Labor Cost

Operation and routine maintenance

D. Parts & Supplier

The operating and maintenance cost is independent of the inlet/outlet sodium levels within the accuracy of this cost study. The estimated operating costs of the 400 gpm (400MW equivalent) fuel treatment system at three utilization rates, two vanadium levels and with two inhibitors are summarized in Table 1.2.2-1. The basis for the elements which make up these total operating and maintenance costs are discussed in detail in the following section on an individual basis.

Table 1.2.2-1

OPERATING AND MAINTENANCE COST OF  
OF FUEL TREATMENT SYSTEM  
400MW COMBINED CYCLE FACILITY

ANNUAL USAGE-HRS.	<u>3000</u>	<u>5000</u>	<u>7000</u>
A. TOTAL ANNUAL COST-\$x10 <sup>6</sup>			
25ppm Vanadium-OS Inhibitor (X)	0.68	1.00	1.30
100 " " " " (X)	1.58	2.50	3.40
" " " WS " (X)	0.68	1.00	1.30
B. UNIT COST- mills/kWh			
25ppm Vanadium-OS Inhibitor (X)	0.57	0.50	0.46
100 " " " " (X)	1.32	1.25	1.21
" " " WS " (X)	0.57	0.50	0.46

(X) OS = oil soluble inhibitor

WS = water-soluble inhibitor

#### 1.2.2.1 Basis For Fuel Treatment System Operating and Maintenance Costs

##### Chemicals

The two chemicals used in a residual fuel treatment system are a chemical demulsifier and a vanadium inhibitor (unless the vanadium is below the threshold level of 0.5 ppm).

Demulsifier- A 400MW system with a fuel flow of 350 gpm, would require an average demulsifier dosage rate of about 1/5000. For the referenced system this amounts to:

4 gph at \$5/gal. = \$20/hr.

Vanadium Inhibitor - A magnesium type vanadium inhibitor must be added at a 3/1 weight ratio of Mg/V. The cost of inhibition depends on the vanadium level in the fuel, the fuel flow rate and the unit cost of the magnesium in the additive.

$$\text{Cost/hr} = 3 \times 10^{-6} \times \text{Fuel Flow} \times \text{Additive Cost}$$

(lbs/hr)      (\$/lb of Mg)

The common denominator for additive cost is the dollars per lb. of magnesium content. There are two common types of magnesium additives; oil-soluble (OS) and water-soluble (WS). The OS is overbased magnesium sulfonate while the WS is magnesium sulfate. The current prices are:

Oil-soluble= 8\$/lb of Mg content

Water soluble= 2\$/lb of Mg content

In the 400MW treatment system the hourly cost at a fuel flow of 350 gpm are:

	<u>25ppm Vanadium</u>	<u>100ppm Vanadium</u>
Oil Soluble	\$100	\$400
Water Soluble	\$ 25	\$100

### Utilities

Water- Potable drinking water or "city water" in the U.S. has a maximum sodium level of about 40ppm. This purity is quite adequate for the fuel washing system. The net water rate has a maximum of 10% of the fuel rate. In the 400 MW system this would be 35 gpm. At an estimated cost of \$0.80/1000 gal, the cost per hour for water would be \$2.

Electricity- Total electrical power consumption of a fuel treatment system for a 400MW power station has been estimated at 550 kWh. At 30 mills/kWh this gives an hourly cost of \$17.

Steam- Steam cost for fuel heating will vary with the fuel viscosity and with the cost assigned to the heating value of the steam. The attached chart 1.2.2-2 shows the fuel temperature required to provide a fuel viscosity of 20 cSt. It also shows the BTU/hr required to heat the fuel to this temperature. Knowing the cost/BTU for the steam and the efficiency of the heat exchanger, the cost of fuel heating can be calculated. For the 400MW power plant example, a fuel with a viscosity of

400 cSt at 100°F was selected as typical. This requires 11.5 million BTU/hr to heat the 350 gpm to 210°F. Using the average reported value for steam cost in the previously referenced EPRI Report AF-1243 of \$1.50 per million BTU, the hourly cost would be \$17.

#### Labor Cost

To operate and perform routine maintenance of a fuel treatment system supplying a base load combined cycle power plant would require one operator per shift around the clock. This translates to five operators working normal work weeks. The average wage for a utility operator in 1980 is \$14/hr. (including 10% for supervision). The annual cost of labor for this system would be \$145,000 total.

#### Parts & Supplies

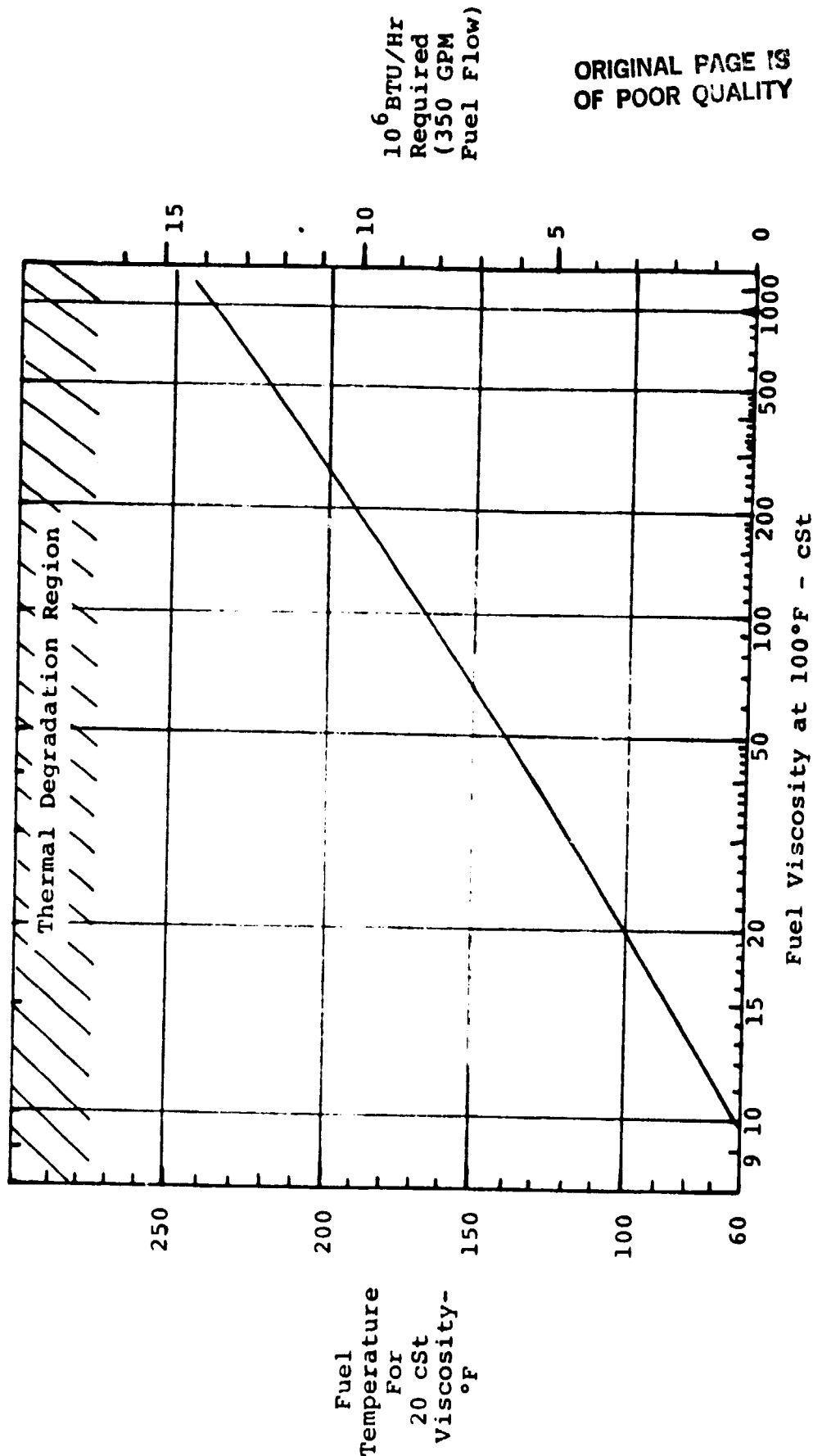
The replacement parts cost for the fuel treatment system hardware has been averaged at 1% of total installed cost. These latter costs have been reported in Table 1.2.1-1 under item B. Using this estimate, the annual parts and supplies cost would range from \$60,000 to \$90,000 with a typical value of \$70,000.

FUEL HEATING REQUIREMENT FOR COMBUSTION

Ave. Fuel Specific Heat - 0.45 BTU/Lb-°F

Ave. Fuel Density - 8.0 Lb/Gal

Reference Fuel Storage Temp. - 60°F





### 1.2.3 Total Residual Fuel Treatment Costs

The total capital investment costs for the selected 400MW combined cycle power plant operating at three sodium contamination levels in the fuel and for three utilization rates (Table 1.2.1-2) and the similar costs for operating and maintenance of the fuel treatment system have been combined in Table 1.2.3-1 to show total fuel treatment costs. Two vanadium levels have been compared using an oil-soluble vanadium inhibitor, and the oil-soluble and water-soluble inhibitors have been compared at a single vanadium level (100 ppm). The 100 ppm of vanadium represents about the maximum level allowable without exceeding typical stack particulate emission regulations. The costs have been rounded off to 0.05 mills/kWh consistent with estimated accuracy of these costs.

Table 1.2.3-1

TOTAL RESIDUAL FUEL TREATMENT COST  
400 MW COMBINED CYCLE FACILITY

(TOTAL FUEL TREATMENT COST = CAPITAL + OPERATION + MAINTENANCE)

VANADIUM (ppm)	INHIBITOR TYPE	SODIUM (ppm)	TOTAL FUEL TREATMENT COST - mills/kWh					
			3000 Hrs/Yr		5000 Hrs/Yr		7000 Hrs/Yr.	
			ED	CD	ED	CD	ED	CD
25	Oil Soluble	0.5	1.55	1.75	1.10	1.25	0.90	1.00
	Oil Soluble	1.0	1.35	1.50	0.95	1.05	0.80	0.85
	Oil Soluble	2.0	1.25	1.35	0.90	1.00	0.75	0.80
100	Oil Soluble	0.5	2.30	2.50	1.85	2.00	1.65	1.75
	Oil Soluble	1.0	2.10	2.25	1.70	1.80	1.55	1.60
	Oil Soluble	2.0	2.00	2.10	1.65	1.75	1.50	1.55
100	Water Soluble	0.5	1.55	1.75	1.10	1.25	0.90	1.00
	Water Soluble	1.0	1.35	1.50	0.95	1.05	0.80	0.85
	Water Soluble	2.0	1.25	1.35	0.90	1.00	0.75	0.80

ED = Electrostatic Desalter  
CD = Centrifugal Desalter

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### 1.3 COMBINED RESIDUAL FUEL TREATMENT AND POWER PLANT OPERATION AND MAINTENANCE

In order to examine the combined effect on incremental costs, the costs of fuel treatment are added to the costs of power plant operation and maintenance. These will demonstrate the combined effect (trade-off) of increased fuel treatment system investment versus the increased parts life (decreased maintenance) resulting from such investment.

Several examples have been chosen to illustrate how this total cost for a Combined Cycle using an electrostatic desalter is affected by the level of treatment versus the cost increments on the power plant.

The effect of additive type (unit cost) is shown by comparing Example 1 with Example 2 and Example 3 with Example 4.

The effect of annual usage rate is shown by comparing Example 1 with Example 3 and Example 2 with Example 4.

Partial Trade-Off Comparing Incremental  
Residual Fuel Treatment Costs and Power  
Plant Operation and Maintenance Costs

Example No. 1

Power Plant	Combined Cycle (400 MW)
Desalting Equipment	Electrostatic
Utilization	3000 hrs/yr.
Annual Write-off	14.6 %/yr.
Type additive	Oil Soluble
Vanadium (ppm)	100

<u>Na in Washed Fuel</u>	<u>Fuel Treatment Cost (mills/kWh)</u>	<u>Power Plant Operation and Maintenance (mills/kWh)</u>	<u>Total</u>
0.5	2.30	2.0	4.30
1.0	2.10	2.4	4.50
2.0	2.00	3.6	5.60

Partial Trade-Off Comparing Incremental  
Residual Fuel Treatment Costs and Power  
Plant Operation and Maintenance Costs

Example No. 2

Power Plant	Combined Cycle (400 MW)
Desalting Equipment	Electrostatic
Utilization	3000 hrs/yr.
Annual Write-off	14.6%/yr.
Type additive	Water Soluble
Vanadium (ppm)	100

<u>Na in Washed Fuel</u>	<u>Fuel Treatment Cost (mills/kwh)</u>	<u>Power Plant Operation and Maintenance (mills/kwh)</u>	<u>Total</u>
0.5	1.55	2.0	3.55
1.0	1.35	2.4	3.75
2.0	1.25	3.6	4.85

Partial Trade-Off Comparing Incremental  
Residual Fuel Treatment Costs and Power  
Plant Operation and Maintenance Costs

Example No. 3

Power Plant	Combined Cycle (400 MW)
Desalting Equipment	Electrostatic
Utilization	7000 hrs.
Annual Write-off	14.6%/yr.
Type additive	Oil Soluble
Vanadium (ppm)	100

<u>Na in Washed Fuel</u>	<u>Fuel Treatment Cost (mills/kWh)</u>	<u>Power Plant Operation and Maintenance (mills/kWh)</u>	<u>Total</u>
0.5	1.65	2.0	3.65
1.0	1.55	2.4	3.95
2.0	1.50	3.6	5.10

Partial Trade-Off Comparing Incremental  
Residual Fuel Treatment Costs and Power  
Plant Operation and Maintenance Costs

Example No. 4

Power Plant	Combined Cycle (400 MW)
Desalting Equipment	Electrostatic
Utilization	7000 hrs/yr.
Annual Write-off	14.6%/yr.
Type additive	Water Soluble
Vanadium (ppm)	100

<u>Na in</u> <u>Washed Fuel</u>	<u>Fuel Treatment</u> <u>Cost (mills/kWh)</u>	<u>Power Plant</u> <u>Operation and</u> <u>Maintenance (mills/kWh)</u>	<u>Total</u> <u>_____</u>
0.5	.90	2.0	2.90
1.0	.80	2.4	3.20
2.0	.75	3.6	4.35

- From an evaluation of these examples it is consistent in every case that increasing the treatment level to reduce the sodium level reduces the combined incremental costs of both fuel treatment and turbine operation and maintenance.



RESIDUAL FUEL TREATMENT COST STUDY  
SIMPLE CYCLE, 30MW POWER PLANT CASE

1.4 ON-SITE RESIDUAL FUEL TREATMENT COSTS

A residual fuel treatment system capable of supplying a specified quality fuel for a 30MW simple cycle gas turbine power plant for intermittent operation (1500 total annual hours) requires a fuel treatment system with a flow rate capacity of 33 gal/min. Only a centrifuge type washing system has been considered because electrostatic desalters lose their cost advantage in small systems, and they are not as adaptable to intermittent operation.

For a gas turbine used 1500 hrs/yr. maximum, the maximum duty cycle has been assumed to be a peak week of 60 hrs.- 12 hrs. per day for 5 days. This is based on utility experience. The fuel treatment system is usually operated continuously from Monday through Friday at a lower fuel flow rate than the turbine with a treated fuel storage tank between the treatment system and the turbine to act as a buffer. Operated in this manner a 25 gpm average fuel flow rate treatment system can satisfy the 40 gpm fuel flow to the gas turbine operated 60 hrs. total/week. This relationship is pictured in Figure 1.4-1. .

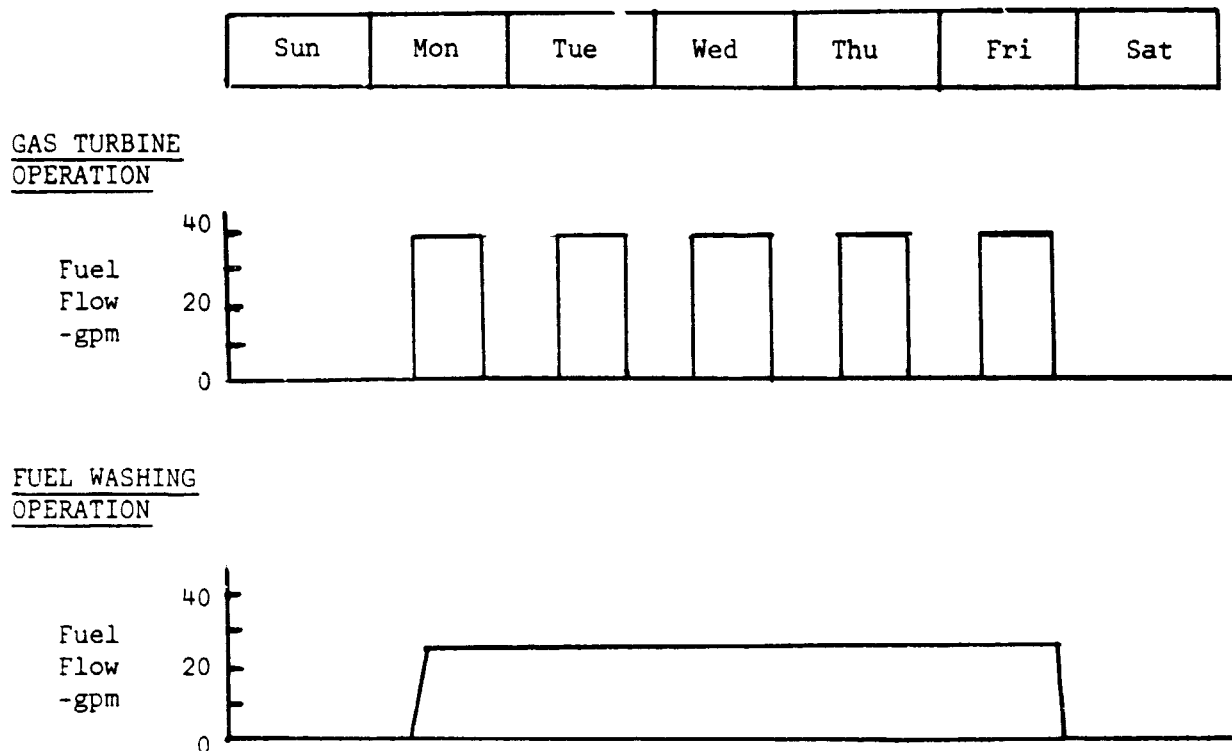
1.4.1 Fuel Treatment System Capital Cost

The 33 gpm fuel treatment system costs have been derived for a reference petroleum residual fuel for three levels of sodium contamination remaining after washing. Table 1.4.1-1 shows the significant properties of the reference fuel and typical installed costs of the systems for the three output sodium levels.

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Figure 1.4-1

TYPICAL DUTY CYCLES OF 30MW SIMPLE CYCLE GAS TURBINE  
AND FUEL WASHING SYSTEM



#### 1.4.1.1 Fuels Outside of Reference Fuel Specifications

For fuels with specific gravity, viscosity and output sodium/input sodium values outside the reference fuel defined in Table 1.4.1-1 multiplication factors can be used for the total installed costs. The reference case with a 1.0 factor is the case shown in Table 1.4.1-3 for a reduction in sodium from 50 ppm in the raw fuel to 1ppm in the washed fuel.

The significant effect of fuel specific gravity, viscosity level and sodium reduction is in the fuel washing equipment. The relative total installed capital costs for fuels other than the reference fuel in Table 1.4.1-1 are listed in Table 1.4.1-3.

Table 1.4.1-1  
RESIDUAL FUEL TREATMENT SYSTEM  
30MW SIMPLE CYCLE POWER PLANT  
CAPITAL COST  
(1980 Dollars)

A. System Description

Fuel Flow Rate -	- 33 gpm max.; 25 gpm typical
Fuel Specific Gravity -	0.96 max.
Fuel Viscosity 100°F -	900 cSt max.
Effluent Water Oil Content -	15 ppm max.
Input Sodium Level -	50 ppm max.

B. Typical Fuel Treatment System Costs

Output Sodium Level, ppm	<u>0.5</u>	<u>1.0</u>	<u>2.0</u>
Total Equipment Cost-\$X10 <sup>6</sup>	1.53	1.33	1.23
Installation Cost-\$X10 <sup>6</sup>	<u>0.67</u>	<u>0.67</u>	<u>0.67</u>
Total Installed Cost-\$X10 <sup>6</sup>	2.20	2.00	1.90

C. Major Equipment Included in Treatment System

Packaged centrifugal desalter system.  
Packaged effluent water cleanup system.  
Inhibitor injection and mixing equipment.  
Residual fuel forwarding and filtration equipment.  
Fuel heaters.  
Fuel analyzer.

Note:

The equipment costs shown are typical values, and are not to be construed as selling prices for any given application. Desalting systems are individually specified for a given application and can vary considerably in cost.

Installation cost at different sites with different systems will also vary from the typical cost shown.

Table 1.4.1-2 shows the investment cost per specific power output for three levels of sodium in the fuel treatment system output.

TABLE 1.4.1-2  
INVESTMENT COST OF FUEL TREATMENT SYSTEM  
30MW SIMPLE CYCLE TURBINE  
(50ppm SODIUM IN INPUT FUEL)  
(1980 Dollars)

<u>Output Sodium in Fuel-ppm</u>	<u>0.5</u>	<u>1.0</u>	<u>2.0</u>
Total Installed Cost-\$10 <sup>6</sup>	2.20	2.00	1.90
* <u>Annual</u> Investment Cost-\$10 <sup>6</sup>	0.321	0.292	0.277
Unit cost-mills/kWh (annual utilization: 1500 hrs.)	7.1	6.5	6.2

\* Based on annual write-off of 14.6% which is currently typical of U.S. utilities.

Table 1.4.1-3

## RELATIVE TOTAL CAPITAL COSTS OF ON-SITE

## FUEL TREATMENT SYSTEM

(33 gpm MAX. FUEL RATE CENTRIFUGE SYSTEM)

Fuel Viscosity	Fuel Specific Gravity	Sodium Out/ Sodium In (ppm)	Total Capital Cost Factor	
Under 900cSt at 100°F	0.96 max. (or blended to 0.96 max.)	2/50	0.95	
		1/50	1.00	
		0.5/50	1.10	
	0.97-0.985 (or blended to 0.97-0.985)	2/20	0.95	
		1/20	1.00	
		0.5/20	1.05	
		2/10	0.90	
		1/10	0.95	
		0.5/10	1.0	
		Over 900 cSt at 100°F	2/50	1.0
			1/50	1.1
			0.5/50	1.2
2/20	1.00			
1/20	1.05			
0.5/20	1.15			
0.97-0.985 (or blended to 0.97-0.985)	2/10	1.00		
	1/10	1.05		
	0.5/10	1.10		

#### 1.4.2 Operating and Maintenance Cost for the Fuel Treatment System

The basis for fuel treatment system operating and maintenance costs were developed for the same reference residual fuel in the 400MW combined cycle case (see Section 1.2.2.1). On the same basis, the operating and maintenance cost per hour of turbine operation have been assigned the following values:

<u>Cost Element</u>	<u>\$/hr. of Turbine Operation*</u>
Chemical Demulsifier (1/5000 of fuel flow)	\$2.5
Vanadium Inhibitor	
Oil soluble-vanadium=25 ppm	12.0
" " " 100 "	48.0
Water Soluble " 100 "	12.0
Water (10% of fuel flow)	0.25
Electricity (75kWh@30 mills/kWh)	2.25
Fuel Heating (Estimated from industry experience)	3.0
System Maintenance (1% of installed cost per year)	10.0
Labor	
(1 man-year total @ \$27,500/yr.)	18

\* 1980 Dollars

For a typical simple cycle turbine operating at 30MW output, the estimated fuel flow rate is 40 gpm. Based on this flow rate and on 1500 hrs. total annual utilization the fuel treatment system operating and maintenance costs become:

TABLE 1.4.2-1

FUEL TREATMENT OPERATING AND MAINTENANCE COSTS

(REFERENCE RESIDUAL FUEL)

(1980 Dollars)

<u>Vanadium- ppm</u>	<u>Inhibitor Type</u>	<u>Operating and Maint. Cost mills/kWh</u>
25	Oil-Soluble	1.60
100	Oil-Soluble	2.80
100	Water Soluble	1.60



#### 1.4.3 Total Residual Fuel Treatment Costs

The total capital investment costs for the 30MW simple cycle power plant operating at three sodium contamination levels in the fuel and for 1500 hrs/yr. and the similar costs for operating and maintenance of the fuel treatment system have been combined in Table 1.4.3-1 to show total fuel treatments costs. Two vanadium levels have been compared using an oil-soluble vanadium inhibitor, and the oil-soluble and water-soluble inhibitors have been compared at a single vanadium level (100 ppm). The 100 ppm of vanadium represents about the maximum level allowable without exceeding typical stack particulate emission regulations. The costs have been rounded off to 0.1 mills/kWh consistent with estimated accuracy of these costs.

TABLE 1.4.3-1

TOTAL FUEL TREATMENT COSTS  
 30MW SIMPLE CYCLE- 1500 hrs/yr  
 (50ppm SODIUM INPUT)  
 (1980 Dollars)

<u>Vanadium ppm</u>	<u>Inhibitor Type</u>	<u>Sodium ppm</u>	<u>Investment Cost mills/kWh</u>	<u>Operating and Maint. Cost mills/kWh</u>	<u>Total Cost mills/kWh</u>
25	Oil Soluble	0.5	7.1	1.6	8.7
		1.0	6.5	1.6	8.1
		2.0	6.2	1.6	7.8
100	Oil Soluble	0.5	7.1	2.8	9.9
		1.0	6.5	2.8	9.3
		2.0	6.2	2.8	9.0
100	Water Soluble	0.5	7.1	1.6	8.7
		1.0	6.5	1.6	8.1
		2.0	6.2	1.6	7.8

### 1.5 Combined Residual Fuel Treatment Cost And Increased Power Plant Operation And Maintenance Costs

In order to examine the combined effect on incremental costs, the costs of fuel treatment are added to the increased costs of power plant operation and maintenance incurred due to residual fuel operation. This demonstrates the trade off of increased investment cost for a fuel treatment system to provide lower sodium levels versus the increased turbine component parts lives (decreased maintenance) resulting from the lower sodium levels.

These combined costs are summarized in Table 1.5-1 for three levels of sodium in the fuel as burned in the turbine, two vanadium levels and for two types of vanadium inhibitors (magnesium additives).

Table 1.5-1

Combined Incremental Cost of Burning Treated  
Residual Fuel

Simple Cycle 30MW gas Turbine

Utilization- 1500 hrs/yr.

(1980 Dollars)

Sodium in Fuel ppm	Vanadium in Fuel ppm	Type of Inhibitor (1)	Increased Cost of Power Plant Oper.&Maint (2) mills/kWh	Fuel Treatment Cost (2) mills/kWh	Total mills/kWh
0.5	25	OS	3.0	8.7	11.7
1.0	"	"	3.4	8.1	11.5
2.0	"	"	4.6	7.8	12.4
0.5	100	OS	3.0	9.9	12.9
1.0	"	"	3.4	9.3	12.7
2.0	"	"	4.6	9.0	13.6
0.5	100	WS	3.0	8.7	11.7
1.0	"	"	3.4	8.1	11.5
2.0	"	"	4.6	7.8	12.4

1) OS = oil-soluble; WS= water soluble

2) Referenced to baseline case of simple cycle operation  
on distillate fuel with 0.5ppm of sodium

3) Reference residual fuel: sodium input to desalter- 50 ppm max.,  
specific gravity - 0.96 max., viscosity- 900 cSt max. at 100°F.

1.6 RESIDUAL FUEL TREATMENT COST STUDY -  
EFFECT OF SYSTEM VARIABLES ON POWER COST

1.6.1 Sensitivity of Cost of Residual Fuel Treatment  
and Turbine Operation and Maintenance to System Variables

These cost studies on petroleum residual fuel for on-site treatment and for increased gas turbine power plant operating and maintenance show certain system variables have greater influence on these costs than others. Table 1.6-1 shows a brief summary of the sensitivity of such costs to power plant type, annual utilization, fuel sodium levels, fuel vanadium level and fuel viscosity/gravity classification.

The greatest effect shown is the difference between the two selected power plant configurations: 400 MW combined cycle and 30 MW simple cycle. The low utilization rate (hrs/yr) tied to the simple cycle case means that the annual investment write-off is distributed over a smaller quantity of power thereby increasing the unit power cost. This is obviously not a controllable variable since the choice between the simple and combined cycle cases is determined by the maximum power requirements and the power-time profile.

The sodium level in the treated fuel burned by the turbine effects the cost of gas turbine operating and maintenance in the 0.5 ppm - 2 ppm range. This can be controlled by the design specifications of the fuel washing system. Except where annual utilization is very low, the lower turbine maintenance cost attainable by low sodium levels (0.5 ppm) outweighs the increased cost of the fuel washing equipment.

The vanadium inhibitor chemical cost is directly proportional to the vanadium level in the fuel. While vanadium cannot be removed by on-site treatment, various levels by refinery processing. (This is covered in Tasks 3 and 4 of this study).

Table 1.6-1

SENSITIVITY OF COSTS OF GAS TURBINE OPERATION AND MAINTENANCE  
AND ON-SITE FUEL TREATMENT TO MAJOR SYSTEM VARIABLES

(1980 DOLLARS)

VARIABLE	RANGE OF VARIABLE	INCREASED MILLS/kWh OVER INDICATED RANGE OF VARIABLE	
		ADDITIONAL POWER PLANT OPER. AND MAINTENANCE	ON-SITE FUEL TREATMENT
1. Power Plant Type	Comb. Cycle, — Simple 3000 hrs/yr Cycle, min. 1500 hrs/ yr max.	1.0	7.0 max.
2. Utilization Rate (Combined Cycle)	7000 — 3000 hrs/yr	None	0.5 max.
3. Sodium level in treated fuel (comb. cycle)	2 — 0.5 ppm	-1.6	0.3 max.
4. Sodium level in delivered fuel	10 — 50 ppm	None	0.2 max.
5. Vanadium level in fuel (comb. cycle)			
a. Oil-soluble inhib- itor	1 — 100 ppm of V	None	1.0 max.
b. water-soluble in- hibitor	1 — 100 ppm of V	None	0.3 max.
6. Fuel Viscosity/Gravity Classification*	Normal — Heavy	None	0.2 max.

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\* Normal = Under 900 cSt @ 100°F/0.96 max. sp. grav.  
Heavy = 900-1200 cSt @ 100°F/0.96-0.985 sp. grav.

The sodium level in the delivered fuel has a small effect on the fuel treatment cost.

A "heavy" fuel requiring a somewhat higher cost fuel washing system also has a small effect on fuel treatment cost. (The case of fuels with viscosities exceeding 1200 cSt at 100°F and/or specific gravities of 0.99 or higher have not been considered. Current practice is to dilute such fuels with cutter stock either by the fuel supplier or at the turbine site and then processing the blend in a normal fashion.)

#### 1.6.2 Effect of Sodium Level in Fuel and Power Plant Configuration on Power Costs

Three cases have been selected to depict the effects of those system variables with the greatest sensitivities on the combined cost of on-site fuel treatment and increased gas turbine power plant operation and maintenance costs. These are shown in Figures 1.6.2-1, 1.6.2-2 and 1.6.2-3 where the unit power cost (mills/kWh) is plotted against the sodium level in the treated fuel burned by the gas turbine. The three cases are:

- Figure 1.6.2-1 - Simple cycle, 30 MW, 1500 hrs/yr
- Figure 1.6.2-2 - Combined cycle, 400 MW, 3000 hrs/yr
- Figure 1.6.2-3 - Combined cycle, 400 MW, 7000 hrs/yr

In the simple cycle 30 MW case, Figure 1.6.3-1, there is a slight minimum in the "total" cost curve at about 1.0 ppm of sodium. This indicates that in a small power plant with a low utilization rate, there is no cost advantage in reducing the sodium level below 1 ppm. There is no such minimum in either combined cycle 400 MW curve indicating that for the large base loaded power plants it pays to reduce the sodium level to as low a value as practicable.

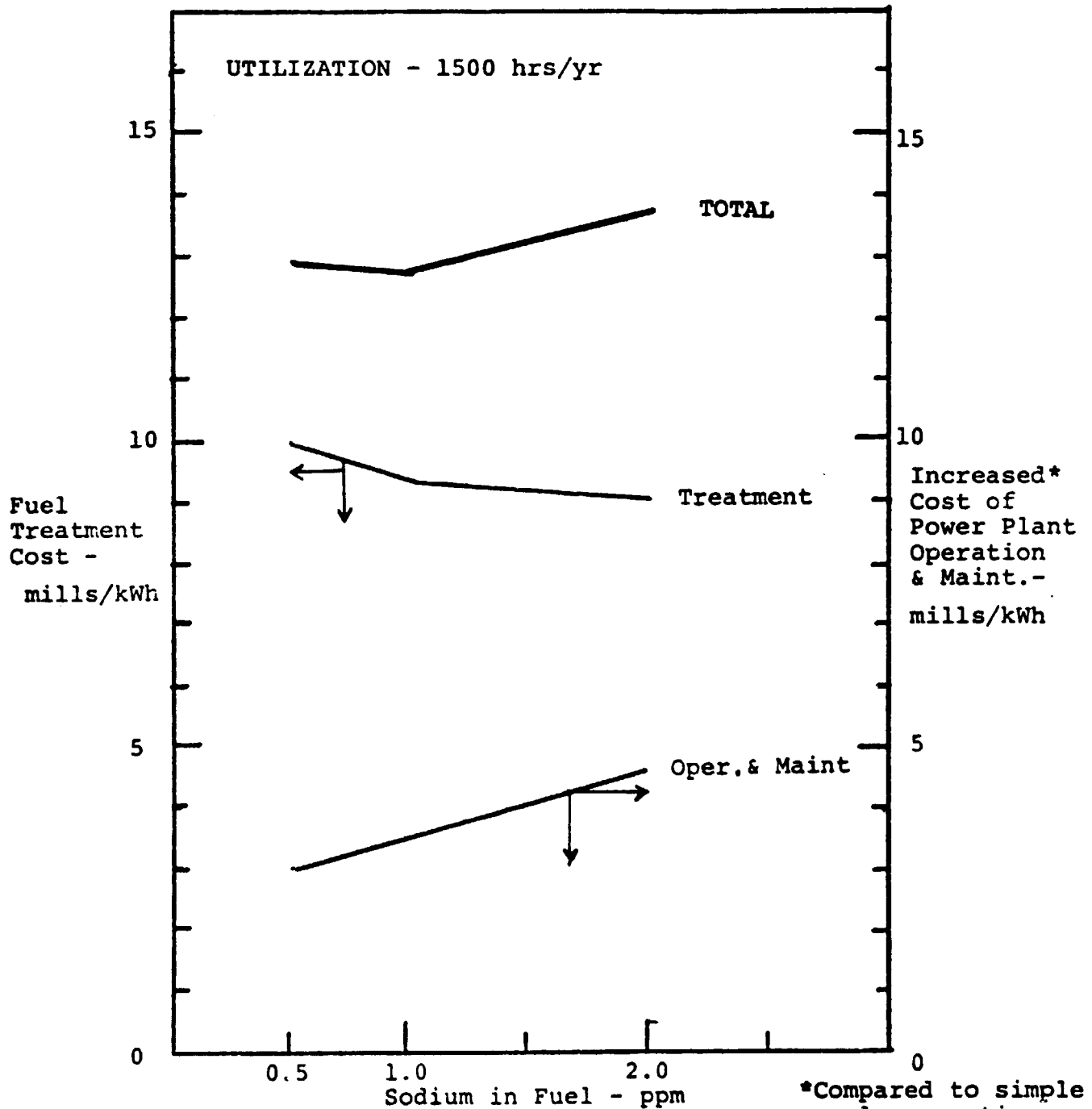
Figure 1.6.2-1

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SIMPLE CYCLE 30MW POWER PLANT  
OPERATING ON RESIDUAL FUEL

Combined Fuel Treatment  
and Increased Operation and Maintenance Costs

Vanadium = 100 ppm  
Oil-soluble Inhibitor



\*Compared to simple  
cycle operation on  
dist. (Na=0.5ppm)



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Figure 1.6.2-2

COMBINED CYCLE 400MW POWER PLANT  
OPERATING ON RESIDUAL FUEL

Combined Fuel Treatment  
and Increased Operation and Maintenance Costs

Vanadium = 100 ppm  
Oil-soluble Inhibitor

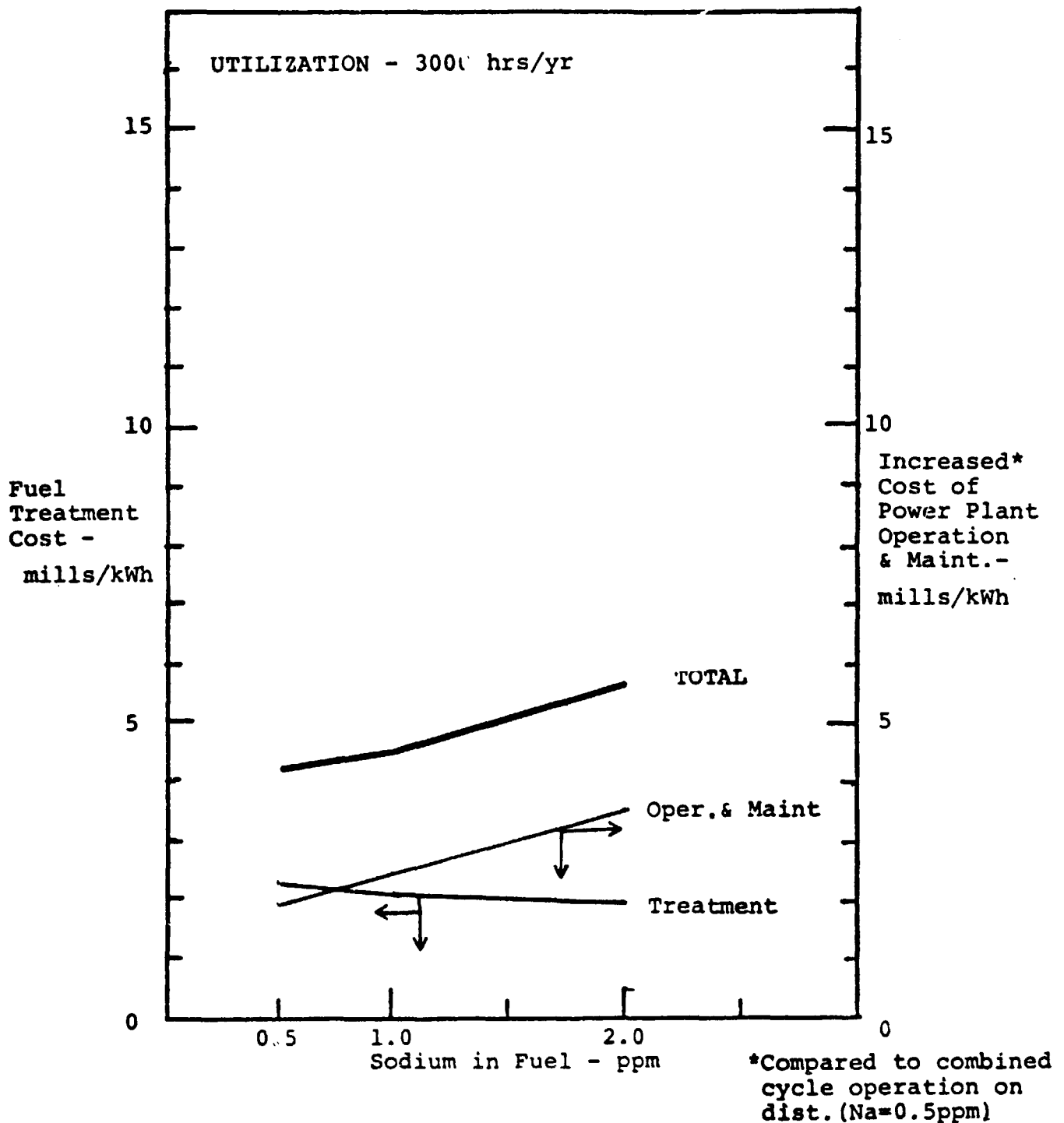


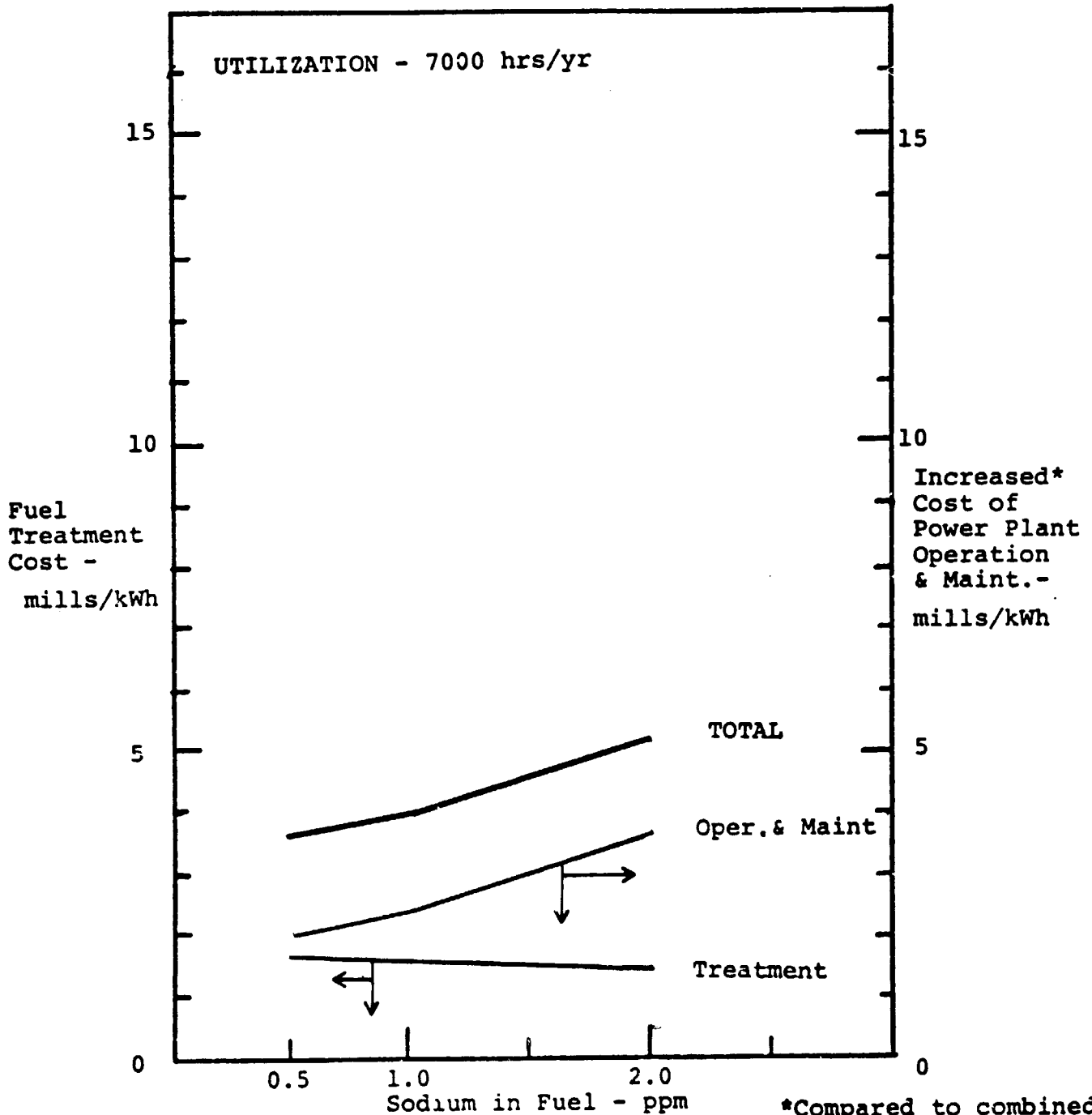
Figure 1.6.2-3

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COMBINED CYCLE 400MW POWER PLANT  
OPERATING ON RESIDUAL FUEL

Combined Fuel Treatment  
and Increased Operation and Maintenance Costs

Vanadium = 100 ppm  
Oil-soluble Inhibitor



\*Compared to combined  
cycle operation on  
dist. (Na=0.5ppm)

## 1.7 RESIDUAL FUEL TREATMENT COST STUDY - CALCULATION OF COSTS OF TREATED RESIDUAL FUEL OPERATION

In Task 5 of this study, economic evaluations will be made of gas turbine power plant operation on petroleum residual fuels with various levels of upgrading, comparing them to baseline operation on clean distillate petroleum fuel. Included in these evaluations will be the following inputs.

1. Cost difference between residual fuel or upgraded residual fuel and clean distillate fuel at the refinery.
- 2a. Cost of on-site fuel treatment to attain a given sodium level and a 3/1 weight ratio of magnesium/vanadium.
- 2b. Added cost of operating and maintaining the gas turbine power plant on treated residual fuel over and above the cost of operation and maintenance on clean distillate fuel in the same power plant.
3. Cost of exhaust gas cleanup for oxides of sulfur and oxides of nitrogen.

Items 1 and 3 are covered elsewhere in the study.

A few specific examples have already been presented for Items 2a and 2b, cost of on-site fuel treatment and added power plant O&M cost respectively. In order that other cases may be studied based on the same cost elements but with different fuel properties, a series of charts have been prepared to facilitate these calculations. These charts allow cost calculations with variations in:

- Sodium level in fuel as received
- Sodium level in washed fuel
- Vanadium level in fuel
- Generic type of vanadium inhibitor
- Fuel viscosity and gravity
- Annual power plant utilization rate -  
combined cycle (3 levels)

Chart A combines the total cost of on-site fuel treatment (investment plus operation and maintenance) with the added cost of power plant operation and maintenance on residual fuels.

The fuel treatment costs for Chart A come from either Chart B (combined cycle power plant) or Chart C (simple cycle power plant).

Chart A

Combined Added Power Plant  
Operating and Maintenance (O&M) Costs  
and Fuel Treatment Costs

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Petroleum Residual Fuel Operation

(All Costs in 1980 Dollars)

Power Plant Type and Size	Sodium Level in Treated Fuel - PPM	Power Plant Added O&M Cost (Note 1)- Mills/kWh	Fuel Treat- ment Cost (Invest plus O&M) (Note 2)- Mills/kWh	Combined Power Plant Added O&M Cost Plus Fuel Treat- ment Cost- Mills/kWh
Comb. Cycle-- 400 MW	0.5	2.0		
	1.0	2.4		
	2.0	3.6		
Simple Cycle-- 30 MW	0.5	3.0		
	1.0	3.4		
	2.0	4.6		

Note 1 - Added cost for using residual fuel over and above the cost of using distillate fuel with 0.5 ppm of sodium in same power plant.

Note 2 - Calculated from Chart B (combined cycle) or Chart C (simple cycle).

CHART B

PETROLEUM RESIDUAL FUEL TREATMENT COSTS  
COMBINED CYCLE - 400 MW POWER PLANT  
3000 - 7000 HRS/YEAR UTILIZATION

(ALL COSTS IN 1980 DOLLARS)

FUEL TREATMENT SYSTEM INVESTMENT COST (14.6% ANNUAL WRITE-OFF)				
A	B	C	D	E
POWER PLANT USE RATE - HRS/YR	TYPE FUEL WASHING SYSTEM (NOTE 1)	INSTALLED SYSTEM COST FOR REFER- ENCE FUEL - MILLS/KWH	FUEL MULT. FACTOR FROM CHART B-1 (NOTE 2)	ADJUSTED INSTALLED SYSTEM COST MILLS/KWH
3000	ED	0.78		
	CD	0.78		
		0.91		
		0.91		
5000	ED	0.47		
	CD	0.47		
		0.55		
		0.55		
7000	ED	0.33		
	CD	0.33		
		0.39		
		0.39		

FUEL TREATMENT SYSTEM OPERATING AND MAINTENANCE (O&M) COSTS					
F	G	H	I	J	K
TYPE OF VANADIUM INHIBITOR (NOTE 3)	UNIT COST OF INHIBITOR MILLS/KWH PER PPM OF V	LEVEL OF VANADIUM IN FUEL - PPM	INHIBITOR COST - MILLS/KWH	ALL OTHER FUEL TREATMENT OPER. & MAINT. COSTS MILLS/KWH	SUB-TOTAL OPER. & MAINT. - COSTS - MILLS/KWH
OS	0.010			0.32	
WS	0.0025			0.32	
OS	0.010			0.32	
WS	0.0025			0.32	
OS	0.010			0.25	
WS	0.0025			0.25	
OS	0.010			0.25	
WS	0.0025			0.25	
OS	0.010			0.21	
WS	0.0025			0.21	
OS	0.010			0.21	
WS	0.0025			0.21	

TOTAL FUEL TREATMENT COST	L (E+K) TOTAL INSTALLED INVEST. COST PLUS OPER. & MAINT. COST - MILLS/KWH
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NOTE 1 - ED = ELECTROSTATIC DESALTER; CD = CENTRIFUGE DESALTER

NOTE 2 - SEE CHART B-1 FOR COST MULTIPLICATION FACTOR BASED ON  
FUEL VISCOSITY, GRAVITY AND INPUT/OUTPUT SODIUM LEVELS

NOTE 3 - OS = OIL-SOLUBLE INHIBITOR; WS = WATER-SOLUBLE INHIBITOR

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Chart B-1

Relative Total Investment Costs of  
On-Site Fuel Treatment Systems

Combined Cycle 400 MW Power Plant  
(400 gpm Max Fuel Rate System)

<u>Fuel Viscosity</u>	<u>Fuel Specific Gravity</u>	<u>Sodium Out/Sodium In (ppm)</u>	<u>Total Investment Cost Multiplication Factor</u>
Under 900 cSt at 100°F	0.96 max. (or blended to 0.96 max.)	2/50	0.90
		1/50	1.00
		0.5/50	1.20
		2/20	0.90
		1/20	0.95
		0.5/20	1.10
		2/10	0.90
		1/10	0.95
		0.5/10	1.0
Over 900 cSt at 100°F	0.97 - 0.985 (or blended to 0.97 - 0.985)	2/50	1.00
		1/50	1.20
		0.5/50	1.40
		2/20	1.00
		1/20	1.10
		0.5/20	1.30
		2/10	1.00
		1/10	1.10
		0.5/10	1.20

CHART C

PETROLEUM RESIDUAL FUEL TREATMENT COSTS  
SIMPLE CYCLE - 30 MW POWER PLANT  
1500 HRS/YR UTILIZATION

(ALL COSTS IN 1980 DOLLARS)

FUEL TREATMENT SYSTEM INVESTMENT COST (14.6% ANNUAL WRITE-OFF)						FUEL TREATMENT SYSTEM OPERATING AND MAINTENANCE (O&M) COSTS					TOTAL FUEL TREATMENT COST
A	B	C	D	E	F	G	H	I	J	K	L
POWER PLANT USE RATE - HRS/YR	TYPE FUEL WASHING SYSTEM (NOTE 1)	INSTALLED SYSTEM COST FOR REFER- ENCE FUEL - MILLS/kWh	FUEL MULT. FACTOR FROM CHART C-1 (NOTE 2)	ADJUSTED CAPITAL COST - MILLS/kWh	TYPE OF VANADIUM INHIBITOR (NOTE 3)	UNIT COST OF INHIBITOR MILLS/kWh PER PPM OF '1	LEVEL OF VANADIUM IN FUEL - PPM	INHIBITOR COST - MILLS/kWh	ALL OTHER FUEL TREATMENT OPER. & MAINT. COSTS MILLS/kWh	SUB-TOTAL OPER. & MAINT. COSTS - MILLS/kWh	TOTAL INVEST. COST PLUS OPER. & MAINT. COST - MILLS/kWh
1500	CD	6.5			OS WS	0.016 0.004			1.20 1.20		

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- NOTE 1 - CD = CENTRIFUGAL DESALTER SYSTEM  
NOTE 2 - SEE CHART C-1 FOR COST MULTIPLICATION FACTOR BASED ON FUEL  
VISCOSITY, GRAVITY AND INPUT/OUTPUT SODIUM LEVELS  
NOTE 3 - OS = OIL-SOLUBLE INHIBITOR; WS = WATER-SOLUBLE INHIBITOR



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Chart C-1

Relative Total Investment Costs of  
On-Site Fuel Treatment Systems

Simple Cycle 30 MW Power Plant  
(33 gpm Max Fuel Rate System)

<u>Fuel Viscosity</u>	<u>Fuel Specific Gravity</u>	<u>Sodium Out/Sodium In (ppm)</u>	<u>Total Investment Cost Multiplication Factor</u>
Under 900 cSt at 100°F	0.96 max. (or blended to 0.96 max.)	2/50	0.95
		1/50	1.00
		0.5/50	1.10
		2/20	0.95
		1/20	1.00
		0.5/20	1.05
		2/10	0.90
		1/10	0.95
		0.5/10	1.0
Over 900 cSt at 100°F	0.97 - 0.985 (or blended to 0.97 - 0.985)	2/50	1.0
		1/50	1.1
		0.5/50	1.2
		2/20	1.00
		1/20	1.05
		0.5/20	1.15
		2/10	1.00
		1/10	1.05
		0.5/10	1.10

1.8 ON-SITE FUEL CLEANUP-HIGH SPECIFIC GRAVITY  
RESIDUAL FUELS

1.8.1 Background

Experience with burning washed residual petroleum fuels has been limited to those fuels with specific gravities less than 0.985 (API greater than 12). The problem of washing high specific gravity fuels has been avoided by fuel selection.

Residual fuels available in the USA come from two main sources: imported (~40%) and domestically produced (~60%). Nearly all the imported No. 6 (burner) fuels presently have specific gravities less than 0.97, but some domestically produced No. 6 fuels have specific gravities 0.99 or higher. The following tabulation shows this distribution based on a recent U.S. Dept. of Energy survey.

DISTRIBUTION OF NO. 6 BURNER FUELS  
IN USA BY SPECIFIC GRAVITY

Source: U.S. DOE Document GETC/PPS-79/4  
"Heating Oils, 1979"

SOURCE	PERCENT OF TOTAL VOLUME USED IN USA	PERCENT OF FUELS* IN SPECIFIC GRAVITY RANGE		
		<0.985	0.985-1.02	>1.02
Domestically Produced	~60	75*	20*	5*
Imported (Largely Caribbean)	~40	100		

\* Based on number of fuels, not total volume.

This survey indicates that there are residual fuels produced in the USA which meet ASTM 4-GT gas turbine liquid fuel requirements but which could not be washed (desalted) by state-of-the-art

techniques and equipment. Although there is not direct correlation between specific gravity and sulfur content, most of the high specific gravity No. 6 fuels have high sulfur levels -- 1.0 to 3.0%.

#### 1.8.2 Potential Desalting Techniques for High Specific Gravity Residual Fuels

##### 1.8.2.1 Fuels with Specific Gravities Close to 1.0

Residual fuels with specific gravities within 1% of the value for water can not be desalted directly by water washing. Where this problem has arisen in the past, the fuel has been blended by the fuel supplier with "cutter stock" to lower the specific gravity to 0.98 or lower. On-site blending has been proposed, but it has proved better to have the fuel supplier carry out the blending.

At the present time, no technique other than fuel blending has been uncovered to desalt residual fuels in this specific gravity range. For years, an alternative was proposed involving increasing the specific gravity of the wash water phase by the addition of a salt which would not harm the gas turbine system. Magnesium sulfate (epsom salt) was proposed, but it was never used in practice and does not represent a viable solution.

##### 1.8.2.2 Fuels with Specific Gravities of 1.02 or Higher

Only a few No. 6 fuels in the USA have specific gravities greater than 1.02 so that this may never become a significant problem. In addition, petroleum residual fuels with very high specific gravities may have other properties which could limit their gas turbine use. The highly naphthenic crudes which normally produce such high specific gravity residual fuels frequently have chemical components which cause poor demulsibility of water. This not only could leave a high percentage

of water in the desalted fuel, but also it could result in very high levels of oil contamination in the effluent wash water which must be cleaned up. Highly naphthenic, high specific gravity residua characteristically also have high fuel-bound nitrogen contents which might also deter their use in gas turbines.

If a requirement for high specific gravity residual fuel desalting does develop in the future, water washing with centrifugation of the wash water would seem to be an approach with a reasonable chance of success. In such a system the fuel would be the heavier phase and the wash water the lighter phase. This is the reverse of today's fuel washing systems, but centrifuges do exist for other applications where the heavier phase is the desired product. An application development program would have to be carried out to adapt such a centrifuge to the specific requirements of a washing system for high specific gravity residual fuels.

## 1.9 ON-SITE FUEL TREATMENT OF SYNFUELS

Unlike petroleum liquid fuels which have well-established properties, synthetic liquid fuels derived from shale and coal are still in a development phase. Neither the commercial liquid fuel product mix nor the specific fuel properties have been resolved; nor will they be fully resolved until future demonstration plant studies have been performed.

The ash-forming trace element content of future synfuels is one of the key considerations for gas turbine application. Both the specific chemical species present and their concentration ranges are important. Therefore, when this Fuel Quality/Processing Study was formulated, on-site removal of ash-forming, deleterious chemical species was one of the areas specified for evaluation. During the elapsed time period, there have been developments in the liquid synfuels field which would seem to rule out any requirement for on-site fuel cleanup to remove ash-forming trace element contaminants. It now appears with some certainty that the "worst case" liquid fuels from both shale retorting and direct coal liquefaction which might be applied to gas turbines would have had sufficient upgrading to reduce the concentration of any ash-forming elements to levels which can be tolerated without fuel treatment.

Shale Oils - The raw shale oil which is the primary product of above ground retorting processes can contain ash-forming trace elements in significant concentrations:

- Iron - 50-150 ppm
- Arsenic - 20-50 ppm
- Nickel - 2-20 ppm
- Vanadium - under 20 ppm
- Copper - 2 ppm
- Sodium plus potassium - (probably less than 1 ppm)

This raw shale oil will have to be stabilized before it can be shipped in a pipeline, and this first upgrading must be done at the conversion facility. In this upgrading process,

any ash-forming trace elements present in the raw fuel would be reduced to low concentrations which could be tolerated by the gas turbine without any fuel cleanup.

As a matter of record, some limited experimental work has been done on the removal of dispersed mineral solids from raw shale oil. R.N. Lucas, U.S. Patent 3,939,625, claimed over 90% inorganic solids removal from raw shale oil using a combination of wash water, surfactant and electrostatic desalter.

Coal Liquids - Coal liquid fuels made by the indirect route-gasification followed by liquid synthesis-would not have any ash-forming trace elements. Coal liquids made by direct liquefaction would have ash-forming contaminants only if these were carried over mechanically from the ash component of the coal feedstock. Unlike petroleum crude oil and residua which have oil-soluble contaminants, namely vanadium and nickel, coal-derived liquids are free of oil-soluble metal-organic contaminants. (Any vanadium in coal is in an inorganic mineral form). The principal ash-forming elements which could be mechanically carried over to liquid coal-derived fuels would be as follows in decreasing order of magnitude:

- Aluminum
- Silicon
- Iron
- Calcium
- Titanium
- Potassium
- Sodium
- Vanadium
- Lead

Such a trace element combination was present in earlier developmental liquid fuels containing vacuum still bottoms which had been de-ashed by mechanical separation techniques. These elements have also appeared at lower but still significant levels in some heavy vacuum distillates where the simple vacuum distillation equipment used allowed ash carryover by entrainment. The use of properly designed and operated vacuum distillation equipment in a commercial coal liquid fuels plant should reduce the ash content of vacuum distillate (heavy distillate) fuels to the low levels typical of today's petroleum-type heavy distillate fuels.

For this Fuel Quality/Processing Study, it has been judged that only true distillate grade coal-derived liquid fuels would be available for gas turbine application as liquid fuels. Two grades seem probable; a low density (middle distillate) grade and a high density (heavy distillate) grade. In each case the total ash content would probably be less than 10 ppm, comparable to commercial No. 2 distillate petroleum fuels.

Although the commercial shale oils and coal-derived liquid fuels potentially available as future gas turbine fuels should not need on-site fuel cleanup because they are essentially ash-free, it should be recognized that they must be kept clean between the fuel refinery and the gas turbine combustor. Water contamination, especially salt water, must be avoided with those liquid synfuels with specific gravities of about 0.99 and above because conventional settling and mechanical purification techniques will not remove the water. Some heavy liquid synfuels, depending on their source and specific processing history, may also have poor demulsibility or water shedding due to the presence of natural emulsifying agents.

In summary, General Electric concurs with the previously stated Gulf position that coal-derived liquids produced by commercial-scale liquefaction plants will be distillates with ash contents of less than 10 ppm, typical of today's petroleum distillate fuels, and tolerable by gas turbines. Thus, for the purposes of this study, neither technical evaluations nor associated cost estimates for on-site treatment will be made for coal liquids projected for use in gas turbines. Similarly, it is concluded that shale oils shipped from shale conversion facilities will contain ash-forming trace element contaminants at low levels which can be tolerated by gas turbines without on-site treatment.



## 2.0 COSTS ASSOCIATED WITH EXHAUST GAS NO<sub>x</sub> REDUCTION SYSTEMS

### 2.1 Introduction

In Task 1.3 the review of the technology for exhaust gas NO<sub>x</sub> reduction identified the dry Selective Catalytic Reduction (SCR) process as the most economically attractive for potential application to combined cycle systems. Wet processes were also considered but are unattractive because of the need to handle large wet streams and a deficiency in operating experience with the high volume flows required for gas turbine application.

In the SCR process, ammonia is mixed with flue gas containing NO<sub>x</sub> and the resulting mixture is passed over a catalyst bed at a specific temperature. In the presence of the catalyst the NH<sub>3</sub> and NO<sub>x</sub> react to produce harmless N<sub>2</sub> and H<sub>2</sub>O. Because the operating temperature range of developed catalysts is 700-800F, this technology cannot currently be considered for application to simple cycle gas turbines which typically have exhaust temperatures in the 950-1020F range.

Japanese companies are known to be developing catalysts with higher temperature capability, but these developments have not yet progressed beyond the laboratory stage.

In a combined cycle the exhaust gas is cooled in the Heat Recovery Steam Generator (HRSG) as the steam is produced. The optimum SCR catalyst temperature range of 700-800F occurs at an

intermediate point within the HRSG gas path. Thus, application of the SCR process to combined cycles requires modifications to the HRSG so that the catalytic reactor section can be located in the appropriate gas temperature region.

At the conclusion of Task 1.3, three SCR processes were identified as having the lowest costs for conventional coal fired steam plants and as being potentially suitable for application to a combined cycle. Of these three, the Hitachi Zosen process appears to have the best potential for lowest costs when integrated into a combined cycle since it utilizes a fixed-bed, plate type catalyst reactor which can more readily be incorporated into the HRSG design. The catalyst systems of the other two processes (UOP and Kurabo-Kuorca) require separate reactor vessels with significant modifications to the HRSG structure to duct the gas from the HRSG to the DeNOx reactor and back to the HRSG. These modifications are estimated to cost significantly more than the modifications to accommodate the Hitachi Zosen fixed bed, plate catalyst, which can essentially be located within the gas flow area of the HRSG by adding additional height to the HRSG (see Figure 1).

A major concern in the application of the SCR process to a combined cycle which is burning sulfur bearing fuel is the potential for serious corrosion of the HRSG fin tube surfaces by attack from ammonium bisulfate ( $\text{NH}_4\text{HSO}_4$ ). Because the catalyst is located part way through the HRSG gas path, there is a large amount of evaporator and economizer finned tube surface downstream of the reactor section.

Ammonium bisulfate forms from a side reaction of the ammonia with  $\text{SO}_3$  present in the exhaust gas from the combustion reactions of sulfur in the fuel. At the high gas temperatures where it forms,  $\text{NH}_4\text{HSO}_4$  is a non-corrosive gas, but as it cools in the heat exchangers downstream of the reactor section, there is a serious potential for it to condense on the finned tube surfaces as a corrosive liquid.

The exact extent of the potential corrosion is not known at this time, but it is felt that even if small quantities of  $\text{NH}_4\text{HSO}_4$  do condense on the carbon steel fin tube surfaces, corrosion rates would be high resulting in unacceptably short life of the fins and tubes. General Electric currently has an experimental test program planned to gain better understanding of this problem and to identify possible design solutions. Until such modifications that might be necessary have been proven, the application of the SCR process to combined cycles which are burning fuel containing sulfur, cannot be considered available technology. If the fuel being burned contains no sulfur (such as clean natural gas), then there is no formation of  $\text{NH}_4\text{HSO}_4$  and hence no corrosion problem.

Since the CDL's and shale oils considered in this study will contain significant levels of sulfur, the SCR cost estimates for application to a combined cycle provided in this section must be considered preliminary and do not include any allowances for design modifications to the HRSG to prevent or reduce  $\text{NH}_4\text{HSO}_4$  corrosion of the fin tubes. Whatever additional costs are required to achieve satisfactory fin tube life in the  $\text{NH}_4\text{HSO}_4$  environment will be additive to the capital cost estimates included here.

## 2.2 Basis for Combined Cycle SCR Cost Estimates

The three principal cost elements of the SCR system are

- a) catalyst material;
- b) catalyst containment, support and necessary HRSG modifications;
- c) ammonia and ammonia storage and distribution systems.

The basis for the cost estimates for items a) and c) presented in this report are the estimates prepared for a 500 MW coal fired steam plant identified in Ref. 1. The basis for the adjustments necessary to adapt the costs for a 400 MW combined cycle are given as follows.

### a) Catalyst

The efficiency of the NO<sub>x</sub> reduction reaction is principally a function of gas temperature and the ratio of gas volume flow to catalyst surface area. The more surface area per unit of gas flow, the more efficient the process.

Since the estimates prepared in Ref. 1 were for a 90% efficient reactor, the same efficiency level was assumed

for the combined cycle. This requires the same ratio of gas flow to catalyst area be maintained, so the catalyst costs for the combined cycle were obtained using the ratio of 400 MW combined cycle exhaust flow ( $8.4 \times 10^6$  lb/hr.) to 500 MW steam boiler flue gas flow ( $4.5 \times 10^6$  lb/hr.).

Since the catalyst has a relatively short life, its costs are considered direct operating expense and not as depreciable capital investment. The catalyst life as guaranteed in Reference 1 is 1 year (7000 operating hours) but actual life of 14,000 hours may be reasonably expected.

b) Catalyst Containment and HRSG Modifications

Cost estimates were prepared for the equipment necessary to contain and support the catalyst material at the appropriate location in the HRSG. Estimates were also prepared for the additional cost involved in splitting the HRSG heat exchanger surfaces to allow insertion of the catalyst section at the correct temperature position.

Factors for installation materials and labor were added and factors were added for A/E fees, contingency and interest during construction consistent with assumptions used in the Ref. 1 report.

c) Ammonia and Ammonia Equipment

To achieve the 90% NO<sub>x</sub> reduction level ammonia has to be mixed with the flue gas on an approx. 1:1 mole ratio to

the NOx in the flue gas. The quantity of  $\text{NH}_3$  required was based on the assumption of 450 ppm NOx in the gas turbine exhaust. This represents a max. NOx concentration for a fuel containing 2% FBN. The annual costs of  $\text{NH}_3$  was based on 7,000 hrs./yr. operation and \$150/short ton cost of ammonia.

The equipment required for injection of the  $\text{NH}_3$  includes a compressor for unloading liquid  $\text{NH}_3$  from truck or rail transport and storage tanks for a 30-day supply. Before  $\text{NH}_3$  is injected into the flue gas, it is vaporized in a steam heated vaporizer and mixed with air supplied from a small air blower to form a 5%  $\text{NH}_3$ -in-air mixture. The  $\text{NH}_3$ -air mixture is injected into the exhaust gas through a grid of nozzles to ensure even distribution and mixing of the  $\text{NH}_3$  into the exhaust gas.

The capital costs for the  $\text{NH}_3$  storage and injection equipment were scaled directly from the costs reported in Ref. 1 on the basis of  $\text{NH}_3$  flow rate.

### 2.3 Exhaust Gas DeNOx Cost Estimates

The capital cost to add SCR process equipment with 90% NOx removal capability to a 400 MW combined cycle plant is estimated to be \$13,250,000 equivalent to \$33.10/kw. This is estimated to be approx. 10% of the installed cost of the base power plant costs.

On the basis that the plant is operated for 7,000 hours/year during which time it produces  $2800 \times 10^6$  KW-HRS., and has gas turbine emissions of 450 ppm NOx, the annual cost of ownership and operation of the SCR System is estimated to be \$14,708,000/YR. which is equivalent to 5.25 mills/KW-HR. Approx. 77% of the annual cost is the catalyst replacement cost, which is based on a catalyst life of 7,000 hours. If the actual life achieved should be 14,000 hours, then the annual operating cost would be reduced to \$9,014,000 which is equivalent to 3.22 mills/KW-HR.

Cost estimate data is presented on the following page.

#### REFERENCE

"Preliminary Economic Analysis of NOx Flue Gas Treatment Processes."

EPA-600/7-80-021

EPRI FP-1253

TVA ECDP B-6

J.D. Maxwell, T.A. Burnett and H.L. Faucett

Tennessee Valley Authority, Muscle Shoals, Alabama

## COMBINED CYCLE SCR SYSTEM COST ESTIMATES

### 1. Basis

Nominal plant rating 400 MW  
Exhaust gas flow  $8.4 \times 10^6$  lb/hr.  
Hitachi Zosen SCR system sized for 90% NOx reduction.  
NOx from gas turbine (reactor inlet) 450 ppm.  
NOx from HRSG stack (reactor outlet) 45 ppm.  
7000 HRS./YR. annual operation  
Annual fixed charge rate (FCR) of 14.6% per Ref. 1.

### 2. Capital Cost

Installed equipment costs including catalyst containment, HRSG modifications, NH<sub>3</sub> storage and injection hardware, A/E fees, contingency and interest during construction (excluding cost of catalyst material).

Total Capital Cost	\$13,250,000	\$33.10/KW
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### 3. Annual Operating Costs

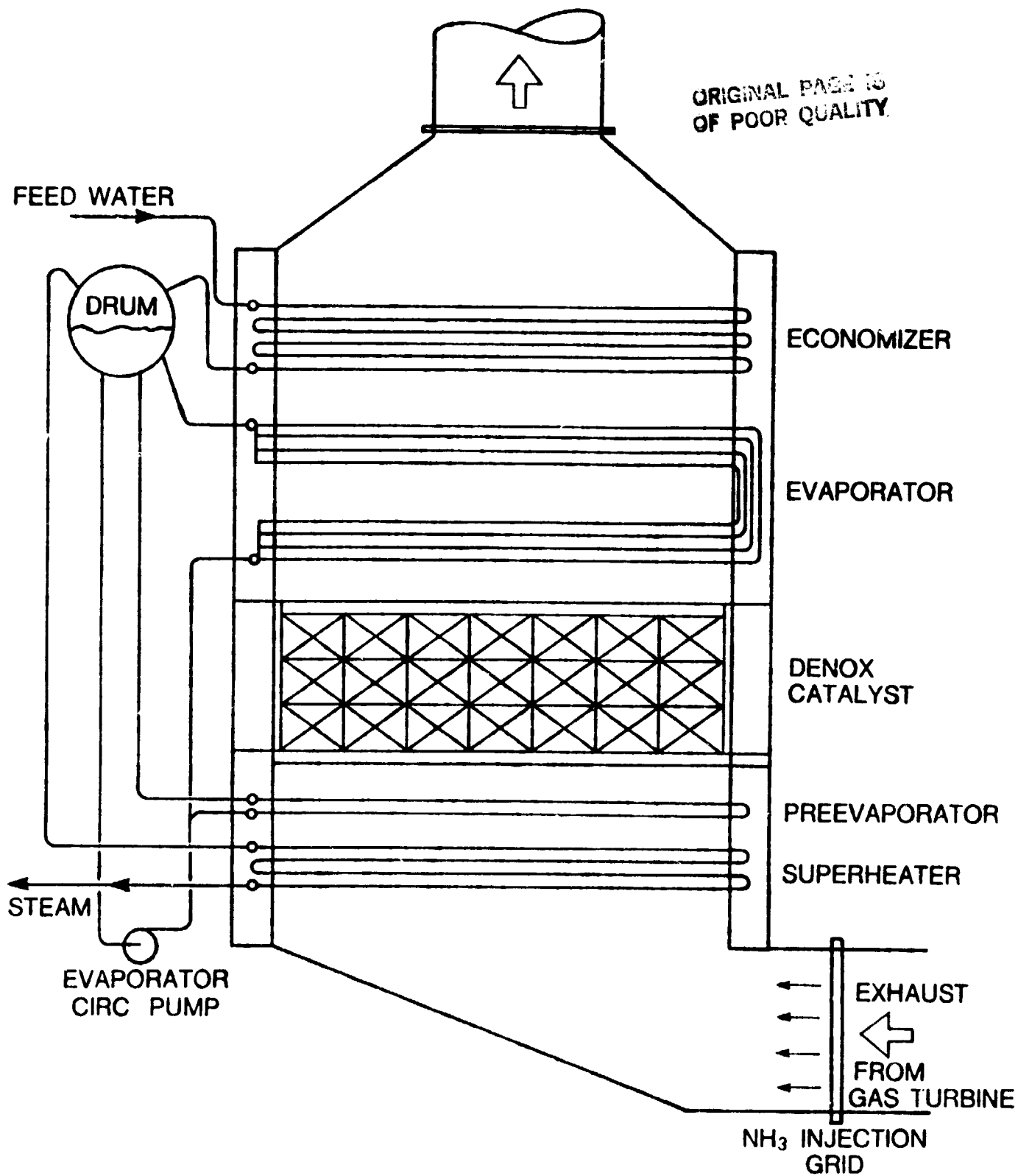
<u>Direct Costs</u>	<u>\$/YR</u>
NH <sub>3</sub> consumption	\$ 1,410,000
Catalyst replacement	11,300,000
Sub-total	<u>\$12,710,000</u>

<u>Indirect Costs</u>	
Capital charges at 14.6% of total capital cost.	\$ 1,903,000
Total Annual Operating Costs	<u>\$14,708,000</u>

for $2800 \times 10^6$ KW-HRS. generated in 1 year.	=	<u>5.25 mills/KW-HR.</u>
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**Figure 1**  
**TYPICAL HRSG ARRANGMENT**  
**WITH DENOX CATALYST**



## 2.4 NOx EMISSION VARIATION AS FUNCTION OF FUEL ORGANIC NITROGEN CONTENT (FBN)

Figure 2 shows the levels of NOx emissions produced by a typical heavy duty, industrial gas turbine equipped with the current state-of-the-art technology combustors.

NOx emissions are produced by two different mechanisms -- thermal NOx which combines atmospheric nitrogen and oxygen, and organic NOx which fixes fuel bound nitrogen with atmospheric oxygen. The production of thermal NOx is highly flame temperature dependent and hence can be controlled (reduced) by the injection of a flame cooling medium such as water or steam. However, the addition of water or steam into the flame zone has the unfortunate effect of increasing the NOx yield from the organic source. Thus the use of steam or water injection as a useful NOx control technology is limited to fuels with quite low FBN levels. To meet the current EPA New Source Performance Standards (NSPS) for gas turbines\* with the maximum amount of steam injection, the upper limit of FBN is 0.1% by weight, and to achieve the Southern California requirements of 140 lb/hr NOx (as NO<sub>2</sub>) (equivalent to approx. 45 ppm) the maximum allowable FBN level is 0.012% by weight.

Since most coal derived liquid fuels and shale oils are expected to contain significantly higher levels of nitrogen, it is clear that steam injection alone will be insufficient to provide a sufficient level of reduction to meet even current levels of emission requirements; and therefore, additional control technology such as SCR will be necessary.

Shown on Figure 2 are the reduction efficiency levels that an SCR system would have to achieve to reduce NOx emission levels to 45 ppmv.

\* EPA New Source Performance Standards for gas turbines.

NSPS is 75 ppmv (dry sample basis) referenced to 15% exhaust oxygen concentration and referenced to a simple cycle heat rate of 13650 Btu/kWhr plus a variable allowance of up to 50 ppm for fuel bound nitrogen levels greater than 0.015%.

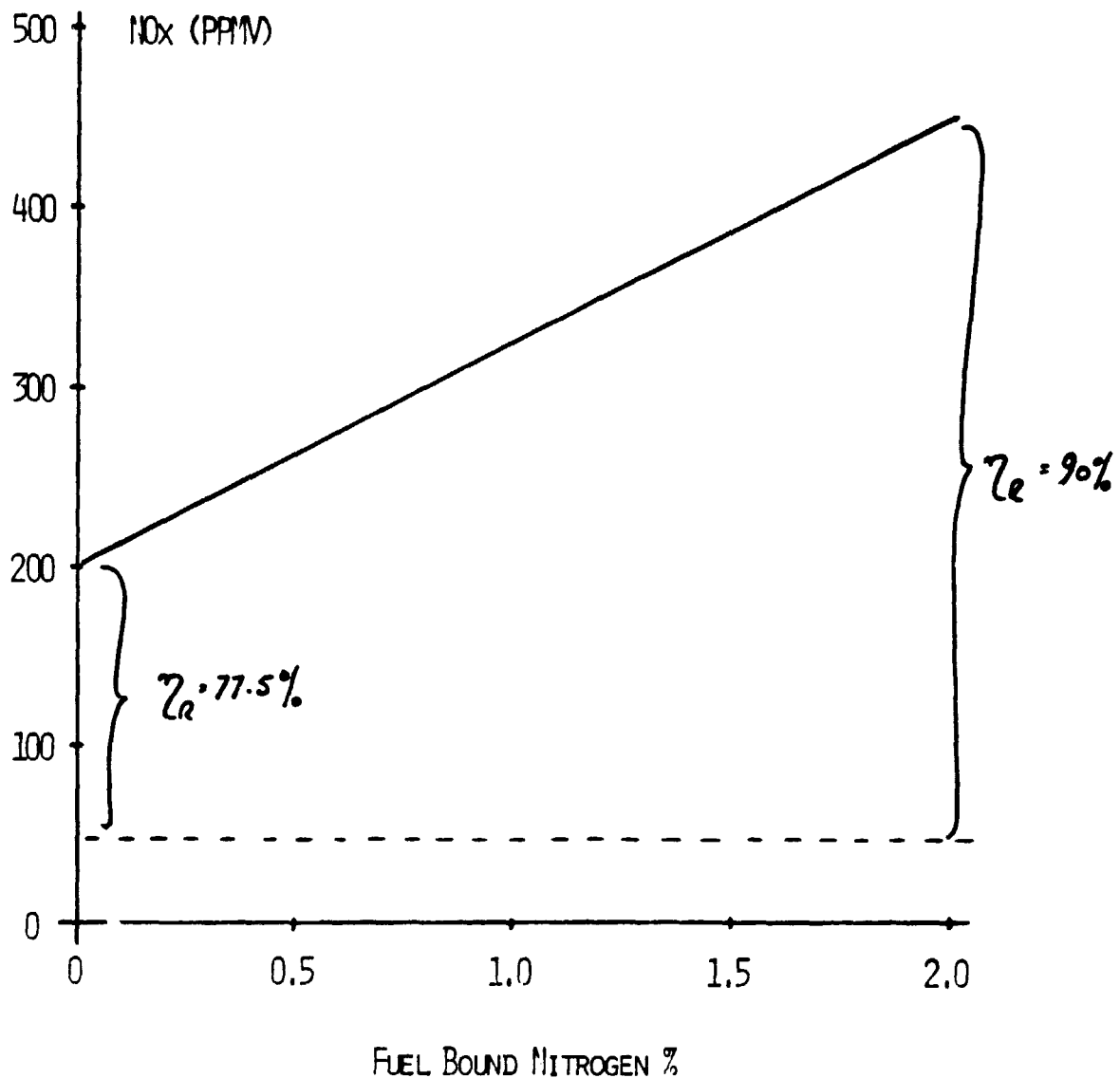
## 2.5 SCR CATALYST REQUIREMENTS AS FUNCTION OF REQUIRED NOx REDUCTION EFFICIENCY

As noted in Section 2.2(a), the efficiency of the NOx reduction process is determined by the ratio of gas flow to catalyst surface area for a specific gas temperature. Since the catalyst cost comprises approx. 50-75% of the total annual operating cost of an SCR system, the required efficiency level will have a strong influence on the total costs. Figure 3 shows the required amount of catalyst relative to a 90% efficient system as a function of NOx reduction efficiency. This curve will be used as the first step in generating overall SCR system costs as a function of NOx removal efficiency.

Figure 2

ORIGINAL PAGE IS  
OF POOR QUALITY

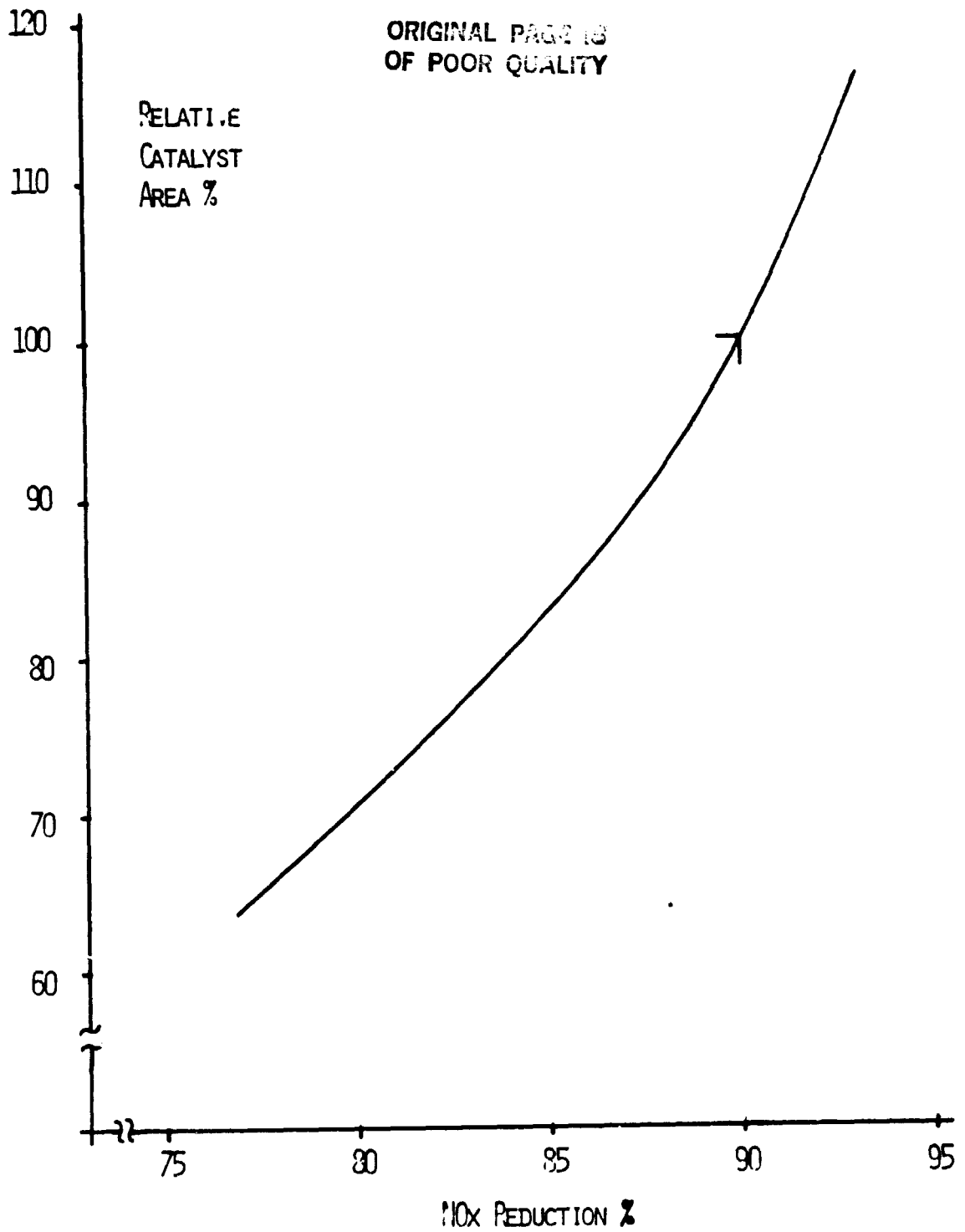
UNCONTROLLED NO<sub>x</sub> EMISSIONS AS FUNCTION  
OF FUEL NITROGEN CONTENT\*



\* FOR CONVENTIONAL HEAVY DUTY GAS TURBINE COMBUSTION TECHNOLOGY.

Figure 3

RELATIVE CATALYST AREA AS FUNCTION OF  
REQUIRED DENOX EFFICIENCY



## 2.6 DeNOx COSTS VS. NOx REMOVAL EFFECTIVENESS

The attached chart summarizes DeNOx system costs for three values of NOx removal effectiveness, 77.5%, 85% and 90%. This range of DeNOx effectiveness provides a constant 45 ppm NOx level at the HRSG stack with gas turbine emissions varying from 200 ppm (0 FBN) to 450 ppm (2.0% FBN).

COMBINED CYCLE SCR SYSTEM COST ESTIMATES

Basis

Nominal plant rating 400 MW.  
7000 HRS/YR. operation 2800 x 10<sup>6</sup> KWHRS. produced.  
Catalyst life 7,000 HRS.  
NOx from HRSG stack (reactor outlet) 45 PPM.

Case #	1	2	3
NOx at reactor inlet PPM	200	300	450
DeNOx reactor efficiency %	77.5	85	90
<u>Capital Costs (\$000's)</u>	12,039	12,523	13,250

Annual Operating Costs (\$/YR. 000's)

Direct Costs

NH <sub>3</sub> Consumption	630	945	1,418
Catalyst replacement	<u>7,400</u>	<u>9,394</u>	<u>11,387</u>

Subtotal direct costs	8,030	10,339	12,805
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Indirect Costs

Capital Charges	1,758	1,828	1,903
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Total Annual Operating Costs	9,788	12,167	14,708
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Annual Costs Mills/KWHR.	3.50	4.34	5.25
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### 3.0 EXHAUST GAS SO<sub>x</sub> CONTROL FOR GAS TURBINES

#### 3.1 Introduction

Orientation economics have been prepared for SO<sub>x</sub> removal from gas turbine exhausts. The 1985 costs are estimated for SO<sub>x</sub> removal from the exhaust gas of a 400 MW combined cycle power plant using a typical liquid fuel. Summary results of the economic evaluation for base load conditions and a fuel sulfur concentration of 2.5 wt% are as follows:

	<u>SO<sub>x</sub> Removal Process</u>		
	<u>Lime/Limestone</u>	<u>Wellman-Lord</u>	<u>Shell-UOP</u>
Total Capital Requirement, \$10 <sup>6</sup>	74	107	96
Operating and Maintenance Costs, \$10 <sup>6</sup> /yr	21	68	9
Exhaust Gas SO <sub>x</sub> Removal Cost, \$/kWh	0.011	0.030	0.008

Exhaust gas SO<sub>x</sub> removal is applicable primarily to a cogeneration or combined cycle power generation system. The most common commercial SO<sub>x</sub> removal process applicable to gas turbines is the lime/limestone process. The Shell-UOP process, with minor modifications, can remove both SO<sub>x</sub> and NO<sub>x</sub> simultaneously and is most economical, but it has not been commercially demonstrated.

The SO<sub>x</sub> removal cost increases significantly as the power plant capacity factor decreases. The effect of fuel sulfur concentration on SO<sub>x</sub> removal cost is nearly the same for both the Shell-UOP and lime/limestone processes, the cost reduction being about 63% of the fuel sulfur reduction.

A survey of the literature on exhaust gas desulfurization, completed under Task II of the Fuels Quality/Processing Study, has shown that desulfurization of the gas turbine exhaust is not currently practiced.<sup>(1)</sup> Gas turbine exhausts generally contain less SO<sub>x</sub> than the maximum permitted by



environmental standards, because the fuel oil is desulfurized during manufacture to minimize turbine corrosion. The current U.S. EPA standard for gas turbines limits the exhaust gas  $\text{SO}_x$  concentration to 150 ppm when converted to 15 percent oxygen on a dry volume basis.<sup>(2)</sup> This limit corresponds to a sulfur content in the fuel of 0.8% by weight, while the current fuel oil specifications for gas turbines usually limit the sulfur content to less than 0.5%.

In the future, with improvements in gas turbine technology and acceptance of high sulfur liquid fuels from petroleum, coal and oil shale,  $\text{SO}_x$  removal from gas turbine exhausts may become necessary. An economic evaluation of representative exhaust gas  $\text{SO}_x$  removal processes is therefore conducted in this study to identify the options specifically suitable for gas turbines.

### 3.2 Exhaust Gas $\text{SO}_x$ Removal Processes

The literature survey<sup>(1)</sup> has shown that of the two types of exhaust gas  $\text{SO}_x$  removal processes, wet and dry, only the wet processes have been developed extensively and commercialized. Of these, the lime/limestone process has been most widely commercialized. Cost data from the lime/limestone process is therefore used as a basis for estimating the economics of a base-case wet desulfurization process for the gas turbines. The economic impact of using another wet process is then assessed by accounting for changes in equipment, chemicals and operating conditions.

Reliable design and cost data for the dry  $\text{SO}_x$  removal processes are not available, because industry experience with these processes is limited to pilot scale operation and conceptual design studies. The Shell-UOP copper oxide process has been selected for economic evaluation in this study, since it is typical of the dry processes applicable to gas turbines and since, with minor modifications, it also can remove  $\text{NO}_x$  simultaneously.

Figure 1 shows a schematic flow diagram of a generalized exhaust gas  $\text{SO}_x$  removal system for a stationary gas turbine. In this system the exhaust gas at  $>1000^\circ\text{F}$  is cooled to  $120^\circ\text{F}$  or  $750^\circ\text{F}$ , the temperature of the wet or dry  $\text{SO}_x$  removal process, respectively, through an energy recovery device. Exhaust gas  $\text{SO}_x$  removal is, therefore, primarily applicable to a combined cycle gas turbine system.

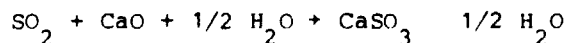
In this study, a 400 MW combined cycle power plant is considered as the basis for evaluating the  $\text{SO}_x$  removal processes. The simple cycle gas turbine, usually in the 30 to 50 MW range, is expected to use fuel desulfurization for  $\text{SO}_x$  emission control and is not considered.

### 3.2.1 Lime/Limestone Process

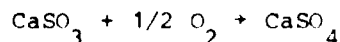
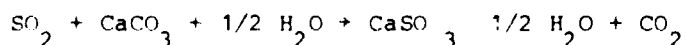
This is the most common commercial process for exhaust gas  $\text{SO}_x$  removal in the United States.<sup>(3,4)</sup> A flow diagram of the process is shown in Figure 2.  $\text{SO}_x$  in the exhaust gas is absorbed by recirculating lime or limestone slurry in a wet scrubber through a series of complex reactions. The products of scrubbing are  $\text{CaSO}_3$  and  $\text{CaSO}_4$  which are separated from the recirculating scrubbing liquid and disposed of as sludge. The cleaned gas leaving the scrubber is reheated to  $\sim 300^\circ\text{F}$  to prevent condensation of the remaining corrosive acids in the ductwork and stack.

The overall  $\text{SO}_x$  absorption reactions are indicated by the following equations:<sup>(5)</sup>

Lime:



Limestone:



Lime is more reactive than limestone, but limestone is cheaper and therefore more widely used.

The equipment required for the lime/limestone process include:

- Crushers and grinders for lime or limestone;
- Slurry preparation, mixing and hold tanks and reactors;
- Wet scrubber for  $\text{SO}_x$  absorption;
- Slurry and liquid pumps, fans, stirrers, piping, ductwork, instrumentation and controls;
- Solids separation, sludge handling and waste disposal units.

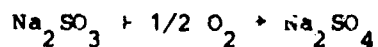
The main advantages of the lime/limestone process are its relative simplicity among wet processes, lower overall costs and utility familiarity based on extensive commercial experience. The disadvantages of this process include a tendency for scaling and plugging of the process internals, need for disposing of large quantities of sludge and greater complexity compared to the dry processes.

### 3.2.2 Wellman-Lord Process

This is a leading wet regenerable process for  $\text{SO}_x$  removal in commercial practice. The  $\text{SO}_x$  absorbing reagent in this process is a solution of sodium sulfite/bisulfite.

Figure 2 shows a flow diagram of the Wellman-Lord process.  $\text{SO}_x$  is removed from the exhaust gas in an absorber. The solution containing the absorbed  $\text{SO}_2$  is sent to an evaporator for regenerating the active reagent. The overall process reactions are represented by the following equations:<sup>(3)</sup>

Absorption:  $\text{SO}_2 + \text{Na}_2\text{SO}_3 + \text{H}_2\text{O} \rightarrow 2 \text{NaHSO}_3$



Evaporation:  $2 \text{NaHSO}_3 \rightarrow \text{SO}_2 + \text{Na}_2\text{SO}_3 + \text{H}_2\text{O}$

Dissolution:  $\text{NaOH} + \text{NaHSO}_3 \rightarrow \text{Na}_2\text{SO}_3 + \text{H}_2\text{O}$

The  $\text{SO}_2$  liberated in the evaporator is used for producing  $\text{H}_2\text{SO}_4$ . In order to prevent buildup of  $\text{Na}_2\text{SO}_4$  in the absorbing liquid, a purge stream is chilled. The crystallized  $\text{Na}_2\text{SO}_4$  is separated and removed from the system. To : sodium loss is made up by adding NaOH to the dissolving tank.

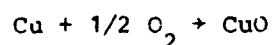
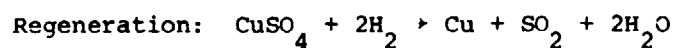
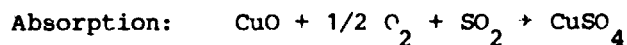
As in the lime/limestone process, the cleaned exhaust gas is reheated to  $\sim 300^\circ\text{F}$ . The equipment used in the Wellman-Lord process includes:

- Wet scrubber for  $\text{SO}_x$  absorption;
- Evaporators, crystallizers and dissolvers for reagent regeneration and makeup;
- Refrigerant and steam supply systems;
- Pumps, fans, stirrers piping and ductwork;
- Solids separation and disposal units.

The main advantages of this process are process reliability, high  $\text{SO}_x$  removal efficiencies (>95%), minimum disposable solid wastes and lack of scaling in the scrubber. The disadvantages include reagent loss through the formation of  $\text{Na}_2\text{SO}_4$ , water pollution from soluble sodium salts, a large steam requirement resulting in derating of the power plant by up to 6% and excessive corrosion of evaporator tubes.

### 3.2.3 Shell-UOP Process

This process is based on the ability of copper oxide to react with  $\text{SO}_2$  and subsequently to be regenerated by hydrogen. The following reactions represent the overall process:



The liberated  $\text{SO}_2$  is either converted to elemental sulfur in a Claus unit or used for producing  $\text{H}_2\text{SO}_4$ .

A block flow diagram of the Shell-UOP process is shown in Figure 4.

The unique feature of the process is a set of specially designed, parallel passage, fixed-bed reactors which use a copper oxide-on-alumina acceptor for  $\text{SO}_2$  removal.<sup>(5)</sup> This configuration is designed to minimize pressure drop and maintain acceptor stability. The process operates at  $750^\circ\text{F}$  for both acceptance of  $\text{SO}_2$  and regeneration of the reactor. Sensible heat from the cleaned gas is recovered by passing it through an air preheater. Natural gas, coal or a liquid fuel is required to produce the hydrogen for regeneration.

The Shell-UOP process has been tested at several power plants in the U.S. and Japan. Reactor stability has been found acceptable in these tests.

The major advantages of this process are minimum solid or liquid waste production and its ability to simultaneously remove  $\text{NO}_x$  and  $\text{SO}_x$  in a single reactor. The disadvantages include process complexity, the need for fuel to produce hydrogen and the relatively inflexible operation of the hydrogen production facility and the Claus plant.<sup>(5)</sup>

### 3.3 Basis of Economics

A 400 MW combined cycle power plant featuring a gas turbine followed by a steam turbine is considered as the basis for evaluating the economics of competing exhaust gas  $\text{SO}_x$  removal processes. The overall energy conversion efficiency of such a system in 1985 is assumed to be 42.7%.<sup>(6)</sup> The typical liquid fuel for the gas turbine is assumed to provide 141,800 Btu/gal.<sup>(7,8)</sup> The power plant exhaust gas containing 15% oxygen by volume under these conditions has a flow rate of  $8.26 \times 10^6$  lb/hr.

The exhaust  $\text{SO}_x$  removal process has to follow the power plant load conditions. Therefore, the annual operating hours considered for evaluating the economics of  $\text{SO}_x$  removal processes are 7000, 5000 and 3000, representing the base, intermediate and peaking loads, respectively.

The sulfur levels of the liquid fuels to be encountered in future gas turbine applications are expected to range from 0.5 to 2.5 wt%. The fuel sulfur concentrations considered in this study are 0.8, 1.5 and 2.5%. In each case the minimum desulfurization requirement is to reduce the exhaust gas  $\text{SO}_x$  concentration to the EPA mandated level of 150 ppm by volume.

Capital, operating and maintenance costs are estimated from published data.<sup>(4,9)</sup> The following escalation factors are used to convert available costs to 1985 levels:

	<u>From</u>	<u>To</u>	<u>Escalation Factor (% per year)</u>
Capital	1976-1980	1980	7.2
	1980	1985	8.0
Operating & Maintenance	1976-1980	1980	8.4
	1980	1985	9.5

Scaleup of process costs from available data to the base plant conditions is done by using an exponential factor of 0.8 in accordance with utility practice.<sup>(9)</sup>

Current utility financing experience indicates that the capital charge rate for estimating annualized costs ranges from 14 to 22%. In this study, a capital charge rate of 14.6% is used<sup>(8)</sup> based on the assumptions shown in Table I.

### 3.3.1 Lime/Limestone Process

Table II shows the 1985 economics of exhaust gas removal by the lime/limestone process. The total capital requirement for the process, applicable to a fuel sulfur concentration of 2.5 wt%, is  $\$74 \times 10^6$ . The operating and maintenance (O&M) costs, at  $\$21 \times 10^6/\text{yr}$ , represent the use of limestone which is the reagent most widely used in the lime/limestone process. The overall cost of exhaust gas removal for the 400 MW combined cycle power plant operating for 7000 hrs/yr is  $\$0.011/\text{kWh}$ .

### 3.3.2 Wellman-Lord Process

Table III shows the 1985 economics of  $\text{SO}_x$  removal by this process, under the same fuel sulfur and power plant operating conditions as those for the lime/limestone process. The total capital requirement for the Wellman-Lord process is  $\$107 \times 10^6$ , nearly 50% more than that for the lime/limestone process. The total O&M costs are  $\$68 \times 10^6/\text{yr}$ , 220% more than the annual O&M costs for the lime/limestone process. These significantly higher costs are caused by process complexity and large utility requirements. The overall cost of  $\text{SO}_x$  removal by the Wellman-Lord process is  $\$0.030/\text{kWh}$ , which is nearly three times the cost for the lime/limestone process.

### 3.3.3 Shell-UOP Process

Table IV shows the 1985  $\text{SO}_x$  removal economics for the Shell-UOP process under conditions similar to those considered for the lime/limestone and Wellman-Lord processes. The total capital requirement for the Shell-UOP

process is  $\$96 \times 10^6$ , which is 30% more than that of the lime/limestone process, because of process complexity. But the total O&M costs at  $\$8 \times 10^6/\text{yr}$  are lower by 60% resulting in an overall  $\text{SO}_x$  removal cost of  $\$0.008/\text{kWh}$ , which is the lowest of the three process costs.

#### 3.3.4 Effect of Capacity Factor on $\text{SO}_x$ Removal Costs

A change in the capacity factor affects primarily the investment based cost per kWh, since the annualized investment cost is fixed. The O&M costs are largely variable and depend directly on the power generated. The effect of capacity factor on the overall  $\text{SO}_x$  removal costs for the three processes considered in this study is shown in Table V and graphically represented in Figure 5. The results show that the cost of  $\text{SO}_x$  removal increases significantly as the capacity factor decreases.

#### 3.3.5 Effect of Fuel Sulfur Concentration on $\text{SO}_x$ Removal Costs

The size of an exhaust gas  $\text{SO}_x$  removal unit depends almost directly on the quantity of sulfur removed. This is because only a part of the exhaust gas proportional to the fuel sulfur concentration requires processing for desulfurization, allowing the rest to bypass the process and be mixed with the cleaned gas before reaching the stack.

The investment required for the process will depend less directly on the fuel sulfur concentration because of the need for additional ductwork, valves and controls required for by passing part of the exhaust gas.

The combination of these effects results in a change of total process investment that may be considered proportional to the design fuel sulfur concentration raised to an exponential factor of 0.8.<sup>(9)</sup>

The variable O&M costs, e.g., chemicals and catalysts, depend directly on fuel sulfur. The fixed O&M costs, e.g., labor and utilities, are assumed to vary with an exponential factor of 0.8, in a manner similar to investment.



The effect of these variations on the cost of exhaust gas  $\text{SO}_x$  removal is shown in Table VI and graphically represented in Figure 6. The results indicate that the overall costs of Shell-UOP and lime/limestone processes decrease by 25-27% by lowering the fuel sulfur concentration from 2.5-1.5%. The effect on the Wellman-Lord process cost is more impressive at 33%, but the total cost of desulfurization by the Wellman-Lord process remains the highest.

Exhaust gas  $\text{SO}_x$  removal plants are generally designed for expected maximum sulfur concentration. The maximum fuel sulfur concentration considered in this study is 2.5 wt%.

Any reduction in fuel sulfur concentration, after the exhaust gas  $\text{SO}_x$  removal plant is constructed and is in operation, affects only the variable operating costs, e.g., steam for releasing absorbed  $\text{SO}_2$ , sulfur recovery, chemicals and waste disposal reagents. The fixed costs that are not affected by the reduction in fuel sulfur concentration include investment-related expenses, utilities related to gas flow control and waste treatment, labor, overhead and catalysts.

The net effects of these changes on the costs of exhaust gas  $\text{SO}_x$  removal by the lime/limestone and the Shell-UOP processes are not significant, as shown in Table VII and Figure 7. The Wellman-Lord process involves major variable operating expenses and therefore is noticeably affected by the reduction in fuel sulfur concentration.

### 3.4 CONCLUSIONS

Gas turbine exhaust gas  $\text{SO}_x$  removal is not currently practiced because the turbine fuel is desulfurized during manufacture to minimize turbine corrosion. Future improvements in gas turbine technology, coupled with acceptance of high sulfur fuels, may make exhaust gas  $\text{SO}_x$  removal necessary for gas turbines.

Exhaust gas  $\text{SO}_x$  removal is applicable primarily to a cogeneration or combined cycle gas turbine/steam turbine system. The simple cycle gas turbine is expected to use fuel desulfurization for  $\text{SO}_x$  emission control.

The most common commercial  $\text{SO}_x$  removal process applicable to gas turbines is the wet lime/limestone process. The Shell-UOP copper oxide process, a dry process, with minor modifications can remove both  $\text{SO}_x$  and  $\text{NO}_x$  simultaneously.

Of the applicable  $\text{SO}_x$  removal processes considered, the Shell-UOP process is most economical, but has not been demonstrated on a base-load commercial scale power plant. The lime/limestone process, which has been extensively commercialized for base-load power plants, costs 40% more for exhaust gas  $\text{SO}_x$  removal than the Shell-UOP process.

The overall cost of exhaust gas  $\text{SO}_x$  removal increases significantly as the capacity factor decreases.

The effect of fuel sulfur concentration on  $\text{SO}_x$  removal cost is nearly the same for both the Shell-UOP and lime/limestone processes. A lowering of fuel sulfur concentration lowers the exhaust gas  $\text{SO}_x$  cost of these processes, the cost reduction being about 63% of the reduction in sulfur concentration.

BNM:WPC (#2471)

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9. Uhl, V. W., "A Standard Procedure for Cost Analysis of Pollution Control Operations; Volume II. Appendices," Report No. EPA-600/8-79-018b, June 1979.
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TABLE I  
 ASSUMPTIONS FOR THE CAPITAL CHARGE RATE (9)

---

	PERCENTAGE OF TOTAL DEPRECIABLE CAPITAL INVESTMENT
YEARS REMAINING LIFE	30
DEPRECIATION-STRAIGHT LINE (BASED ON YEARS REMAINING LIFE OF POWER UNIT)	3.3
INTERIM REPLACEMENTS (EQUIPMENT HAVING LESS THAN 30-YR LIFE)	0.7
INSURANCE AND PROPERTY TAXES	2.0
TOTAL RATE APPLIED TO ORIGINAL INVESTMENT	6.0
	PERCENTAGE OF UNRECOVERED CAPITAL INVESTMENT (A)
COST OF CAPITAL (CAPITAL STRUCTURE ASSUMED TO BE 60% DEBT AND 40% EQUITY)	
BONDS AT 10% INTEREST	6.0
EQUITY (B) AT 14% RETURN TO STOCKHOLDER	5.6
INCOME TAXES (FEDERAL AND STATE) (C)	5.6
TOTAL RATE APPLIED TO DEPRECIATION BASE	17.2 (D)

- 
- (A) ORIGINAL INVESTMENT YET TO BE RECOVERED OR "WRITTEN OFF."  
 (B) CONTAINS RETAINED EARNINGS AND DIVIDENDS.  
 (C) SINCE INCOME TAXES ARE APPROXIMATELY 50% OF GROSS RETURN,  
 THE AMOUNT OF TAXES IS THE SAME AS THE RETURN ON EQUITY.  
 (D) APPLIED ON AN AVERAGE BASIS, THE TOTAL ANNUAL PERCENTAGE  
 OF ORIGINAL FIXED INVESTMENT FOR NEW (30 YR) PLANTS WOULD  
 BE  $6.0\% + 1/2 (17.2\%) = 14.6\%$ .

## TABLE II

LIME/LIMESTONE PROCESS FOR EXHAUST GAS SO<sub>x</sub> REMOVALECONOMICS  
1985

BASIS: 400 MW COMBINED CYCLE POWER PLANT OPERATING 7000 HRS/YR  
 EXHAUST GAS FLOW RATE: 8.26 MILL'N LB/HR  
 OXYGEN IN EXHAUST GAS: 15% BY VOLUME  
 FUEL SULFUR: 2.5% BY WEIGHT

## INVESTMENT, \$ MILL'N

INSTALLED CAPITAL, SCRUBBER, AUXILIARIES, WASTE TREATMENT, INSTRUMENTATION AND CONTROL	52.24
INTEREST DURING CONSTRUCTION	8.83
STARTUP	4.20
	-----
TOTAL DEPRECIABLE INVESTMENT	65.57
LAND	1.57
WORKING CAPITAL	6.55
	-----
TOTAL CAPITAL REQUIREMENT	73.69

## OPERATING &amp; MAINTENANCE (O&amp;M), \$ MILL'N/YR.

CHEMICALS AND MATERIALS	0.92
LABOR, OVERHEAD AND UTILITIES	20.16
	-----
TOTAL O&M	21.08
ANNUALIZED INVESTMENT, 14.6% OF TOTAL CAPITAL REQUIREMENT, \$MILL'N/YR	10.76
	-----
TOTAL ANNUAL EXPENSE, \$MILL'N/YR	31.84
EXHAUST GAS SO <sub>x</sub> REMOVAL COST, \$/KWH	0.011

# TABLE III

## WELLMAN-LORD PROCESS FOR EXHAUST GAS SO<sub>2</sub> REMOVAL

### ECONOMICS 1985

BASIS: 400 MW COMBINED CYCLE POWER PLANT OPERATING 7000 HRS/YR  
EXHAUST GAS FLOW RATE: 8.26 MILL'N LB/HR  
OXYGEN IN EXHAUST GAS: 15% BY VOLUME  
FUEL SULFUR: 2.5% BY WEIGHT

#### INVESTMENT, \$ MILL'N

INSTALLED CAPITAL, SCRUBBER, SULFUR RECOVERY SYSTEM, AUXILIARIES, WASTE TREATMENT, INSTRUMENTATION AND CONTROL	76.18
INTEREST DURING CONSTRUCTION	12.80
STARTUP	6.09
TOTAL DEPRECIABLE INVESTMENT	95.07
LAND	2.28
WORKING CAPITAL	9.50
TOTAL CAPITAL REQUIREMENT	106.65

#### OPERATING & MAINTENANCE (O&M), \$ MILL'N/YR.

CHEMICALS AND MATERIALS	2.94
LABOR, OVERHEAD AND UTILITIES	64.51
TOTAL O&M	67.45
ANNUALIZED INVESTMENT, 14.6% OF TOTAL CAPITAL REQUIREMENT, \$MILL'N/YR	15.60
TOTAL ANNUAL EXPENSE, \$MILL'N/YR	83.05
EXHAUST GAS SO <sub>2</sub> REMOVAL COST, \$/KWH X	0.030

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TABLE IV

SHELL-UOP PROCESS FOR EXHAUST GAS SO<sub>x</sub> REMOVAL

ECONOMICS  
1985

BASIS: 400 MW COMBINED CYCLE POWER PLANT OPERATING 7000 HRS/YR  
EXHAUST GAS FLOW RATE: 8.26 MILL'N LB/HR  
OXYGEN IN EXHAUST GAS: 15% BY VOLUME  
FUEL SULFUR: 2.5% BY WEIGHT

INVESTMENT, \$ MILL'N

INSTALLED CAPITAL, 2 REACTOR/REGENERATOR UNITS, SULFUR RECOVERY SYSTEM, AUXILIARIES, WASTE TREATMENT, INSTRUMENTATION AND CONTROL	69.42
INTEREST DURING CONSTRUCTION	11.67
STARTUP	5.55
TOTAL DEPRECIABLE INVESTMENT	86.64
LAND	2.09
WORKING CAPITAL	6.95
TOTAL CAPITAL REQUIREMENT	95.68

OPERATING & MAINTENANCE (O&M), \$ MILL'N/YR.

CATALYSTS CHEMICALS AND UTILITIES	8.31
LABOR AND OVERHEAD	0.23
TOTAL O&M	8.54
ANNUALIZED INVESTMENT, 14.6% OF TOTAL CAPITAL REQUIREMENT, \$MILL'N/YR	13.97
TOTAL ANNUAL EXPENSE, \$MILL'N/YR	22.51
EXHAUST GAS SO <sub>x</sub> REMOVAL COST, \$/KWH	0.008

TABLE V

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-----

BASIS: 1985 COSTS FOR A 400 MW COMBINED CYCLE POWER PLANT  
 FUEL SULFUR: 2.5% BY WEIGHT  
 EXHAUST GAS FLOW RATE: 8.26 MILL'N LB/HR WITH 15% OXYGEN

PROCESS	CAPACITY FACTOR, HOURS/YEAR	SO <sub>x</sub> REMOVAL COSTS			
		\$ MILL'N/YR			\$/KW.H
		INVESTMENT	O&M	TOTAL	
LIME/LIMESTONE	7,000	10.76	21.08	31.84	0.011
	5,000	10.76	18.72	29.48	0.015
	3,000	10.76	16.37	27.13	0.023
WELLMAN-LORD	7,000	15.60	67.46	83.06	0.030
	5,000	15.60	59.91	75.51	0.038
	3,000	15.60	52.37	67.97	0.057
SHELL-UOP	7,000	13.97	8.55	22.52	0.008
	5,000	13.97	6.17	20.14	0.010
	3,000	13.97	3.79	17.76	0.015



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TABLE VI

EFFECT OF FUEL SULFUR CONCENTRATION

BASIS: 1985 COSTS FOR A 400 MW COMBINED CYCLE POWER PLANT  
OPERATING 7000 HRS./YR.  
EXHAUST GAS FLOW RATE: 8.26 MILLION LB/HR  
OXYGEN IN EXHAUST GAS: 15% BY VOLUME

FUEL SULFUR, WT%	SO <sub>x</sub> REMOVAL COSTS, \$/KW.H		
	LIME-LIMESTONE PROCESS	WELLMAN- LORD PROCESS	SHELL-UOP PROCESS
2.5	0.011	0.030	0.008
1.5	0.008	0.021	0.006

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Table VII

EFFECT OF CHANGES IN FUEL SULFUR CONCENTRATION ON FGD COST

1985

Basis: FGD plant designed for 2.5% S in fuel

Operating hours/yr: 7000

Exhaust gas flow rate: 8.26 mill'n lb/hr

Oxygen in exhaust gas: 15% by volume

Fuel sulfur equivalent of  $SO_x$  emission after  
exhaust gas  $SO_x$  removal: 0.8 wt %

Concentration of S in Fuel wt %	$SO_x$ Removal Cost, \$/kW.h Process		
	Lime/Limestone	Wellman-Lord	Shell-UOP
2.5	0.011	0.030	0.008
1.5	0.010	0.026	0.007
0.8	0.008	0.020	0.005

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FIGURE 1  
GENERALIZED FLOW DIAGRAM OF THE EXHAUST GAS  $\text{SO}_x$  REMOVAL SYSTEM

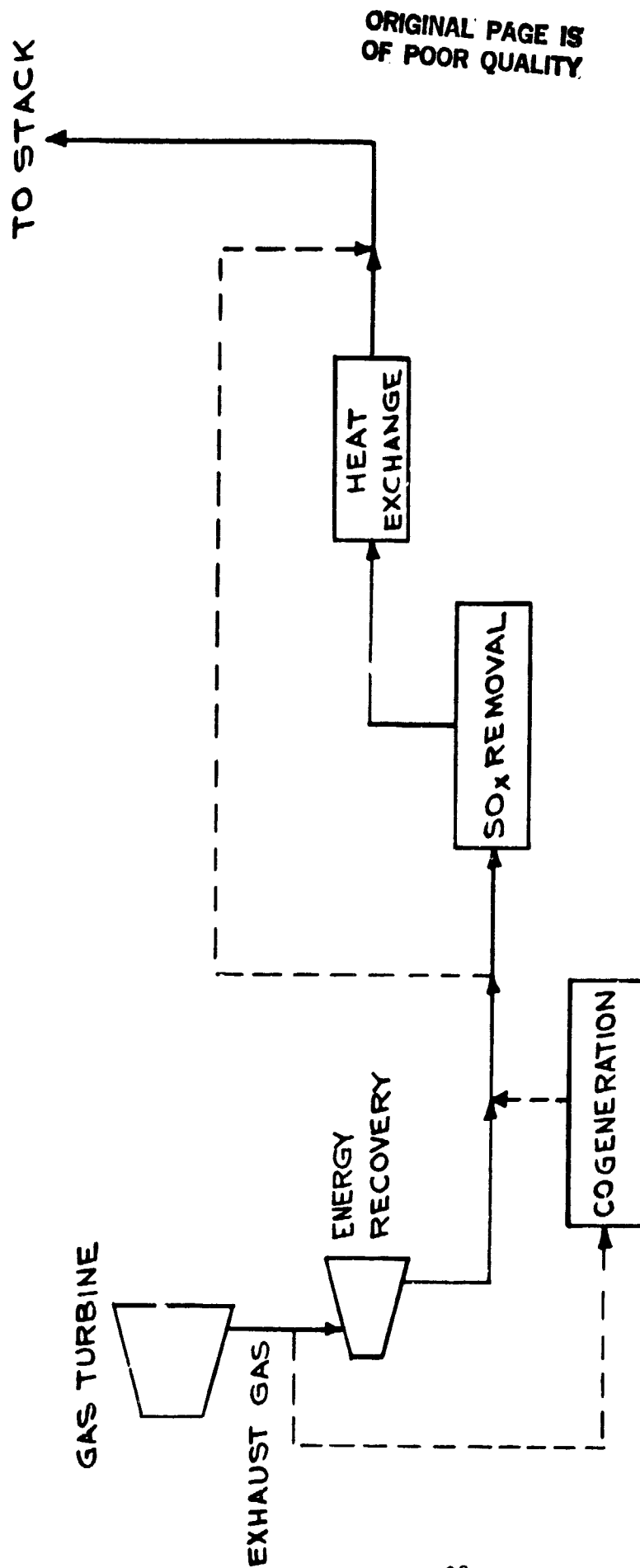
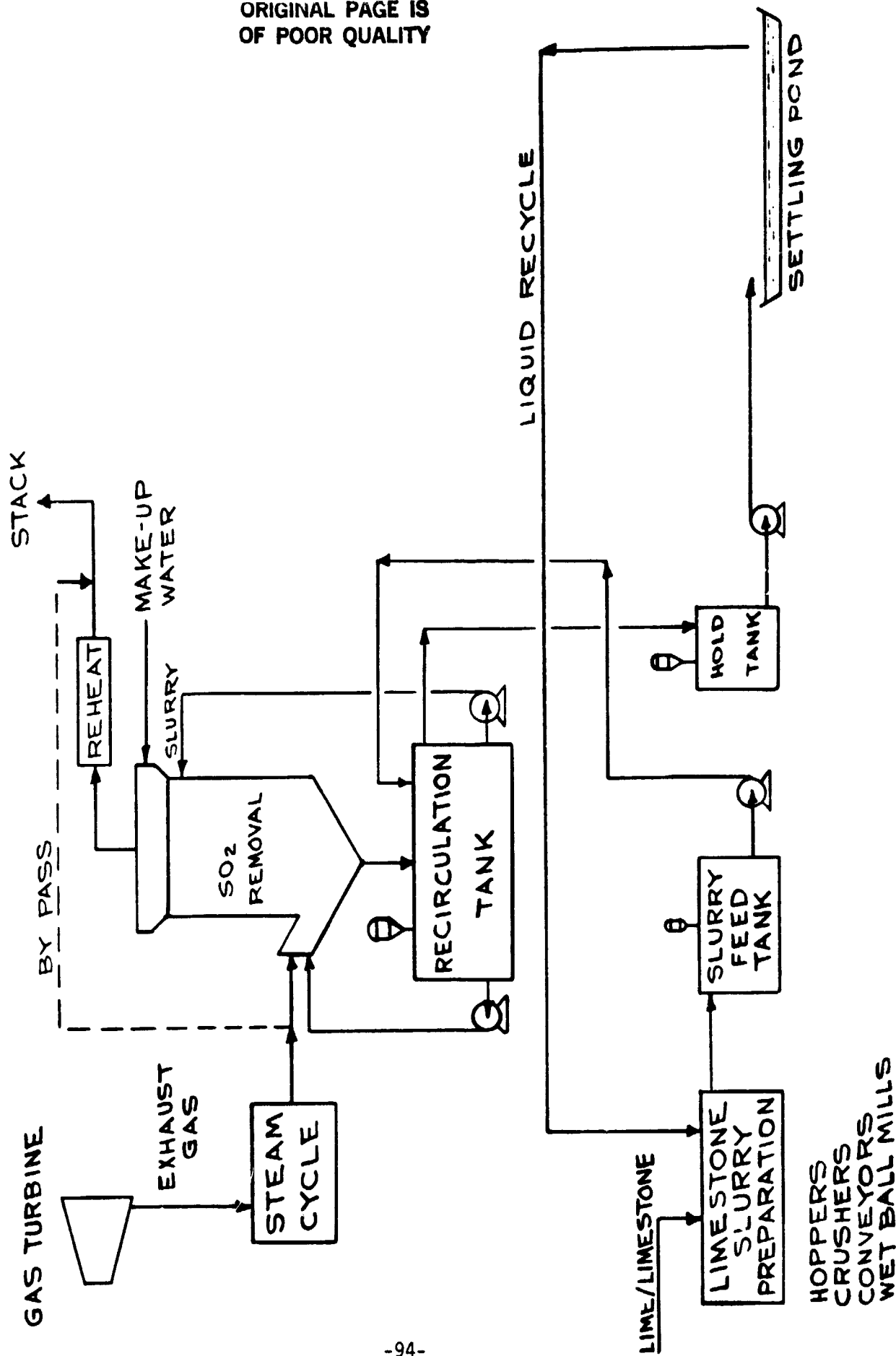


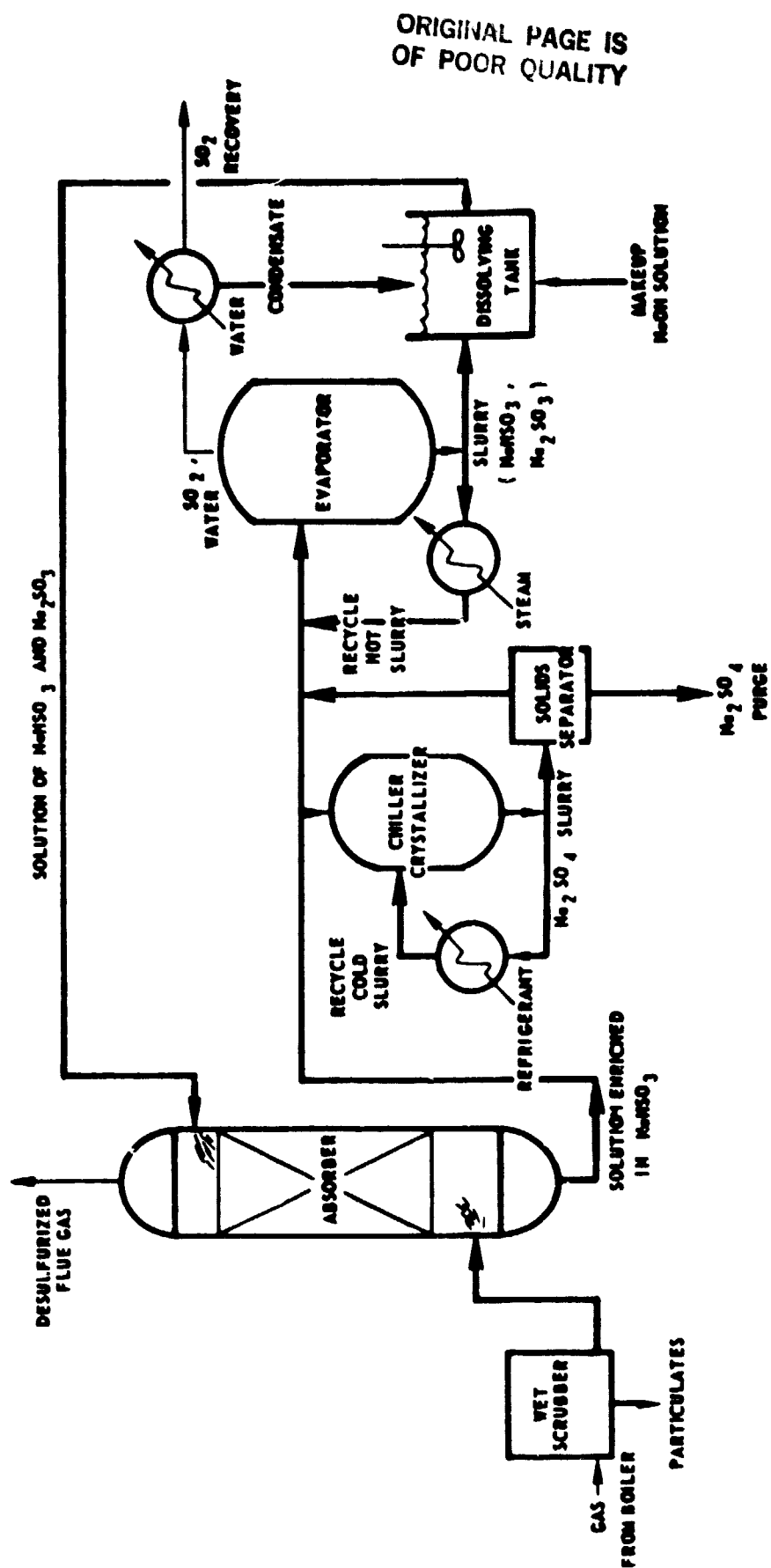
FIGURE 2

EXHAUST GAS SO<sub>2</sub> CONTROL SYSTEM FOR THE BASE CASE:  
400 MW COMBINED CYCLE POWER PLANT, LIME/LIMESTONE WET SCRUBBER

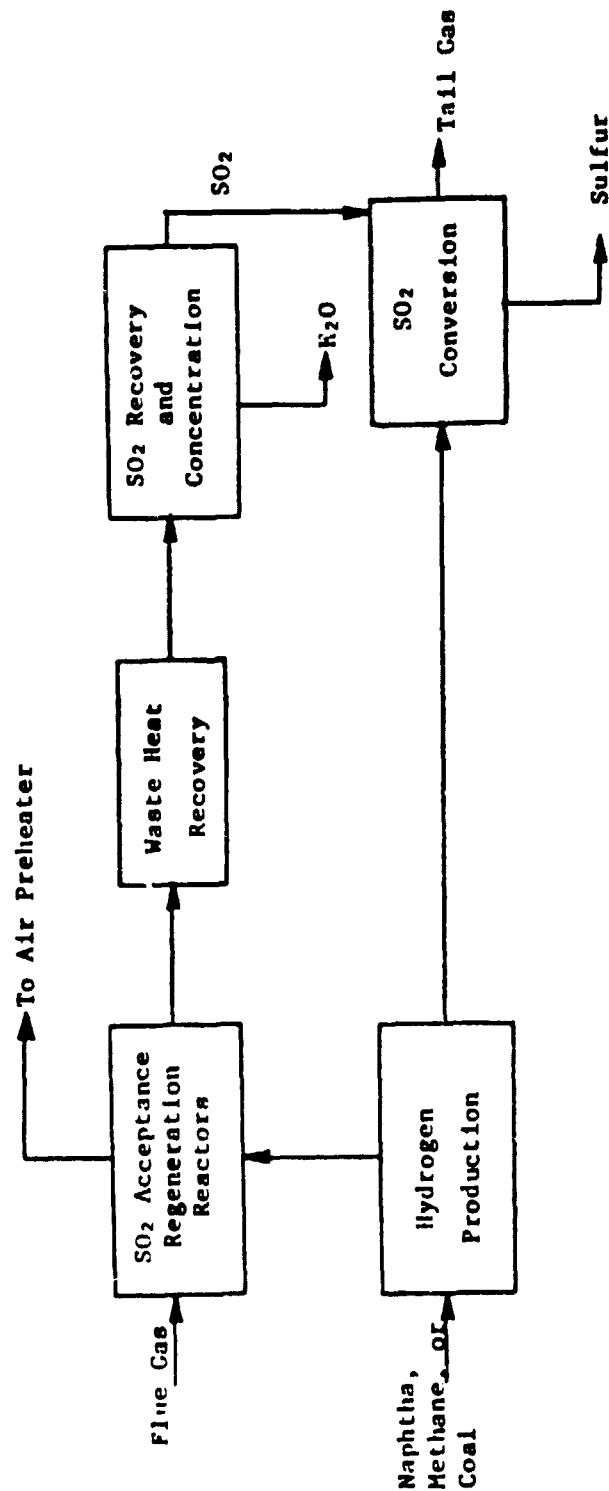


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FIGURE 3  
WELLMAN-LORD PROCESS (4)

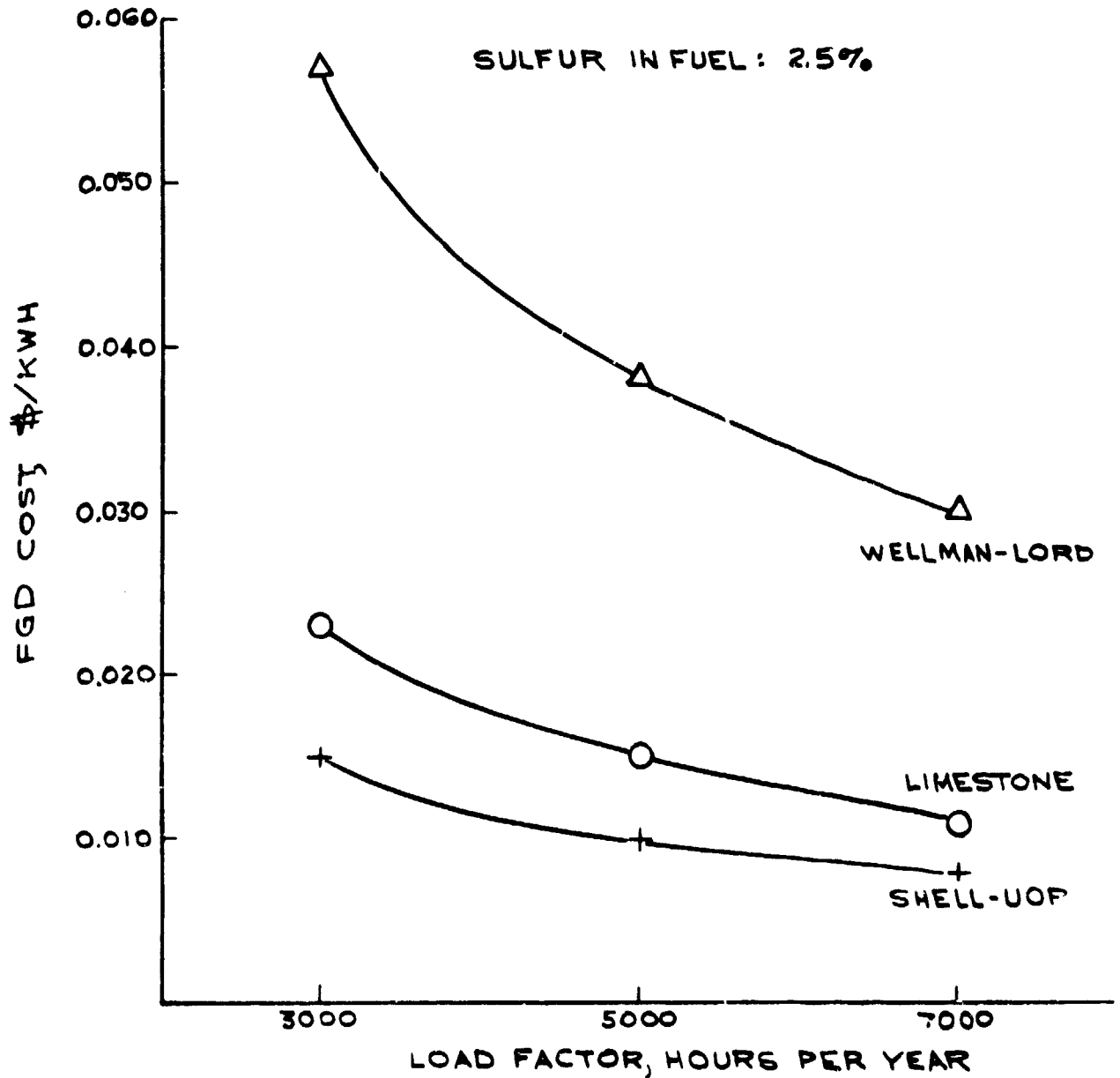


**FIGURE 4**  
**SHELL FLUE GAS DESULFURIZATION PROCESS** (5)



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FIGURE 5  
COMPARISON OF EXHAUST GAS SO<sub>2</sub> REMOVAL  
PROCESSES



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FIGURE 6  
EFFECT OF FUEL SULFUR ON FGD COST

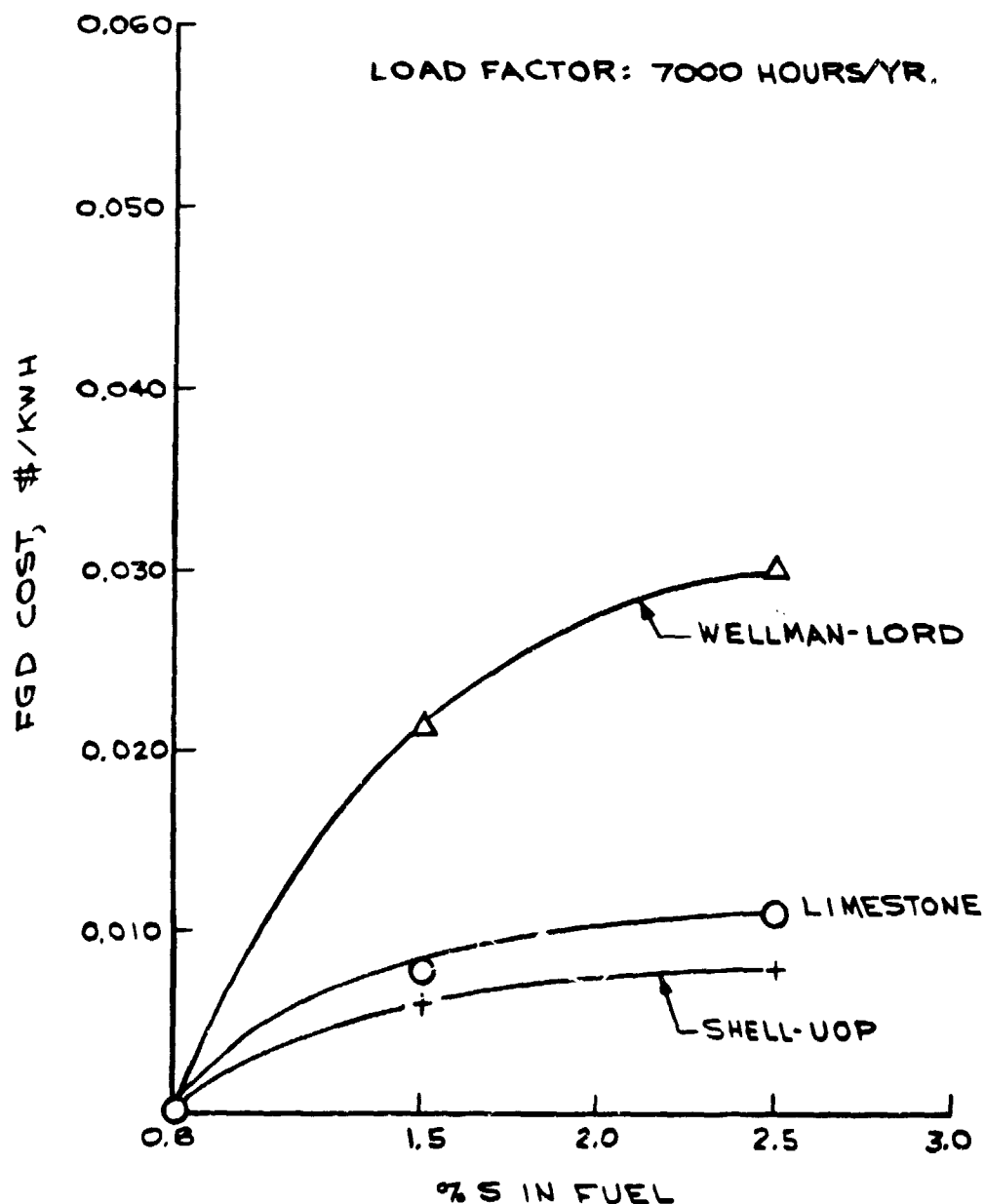




FIGURE 7

EFFECT OF CHANGES IN FUEL SULFUR CONCENTRATION ON FGD COSTS

